PB 203 958

THE HIGH SULFUR COMBUSTOR

A Study of Systems
for
Coal Refuse Processing

VOLUME ONE
(NARRATIVE SUMMARY)

FINAL REPORT
to
DIVISION OF PROCESS CONTROL ENGINEERING
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
ENVIRONMENTAL HEALTH SERVICE

U.S. DEPARTMENT
of
HEALTH, EDUCATION and WELFARE

CONTRACT NO. CPA 22-69-151

CONSULTING DIVISION
CHEMICO
CHEMICAL CONSTRUCTION CORPORATION
NEW YORK, NEW YORK

Reproduced by
NATIONAL TECHNICAL
INFORMATION SERVICE
Springfield, Va. 22151

BIBLIOGRAPHIC DATA SHEET	1. Report No. APTD-0768	2.	3. Recipient's Accession No.

4. Title and Subtitle
The High Sulfur Combustor
A Study of Systems for Coal Refuse Processing
Volume One (Narrative Summary)

5. Report Date
February 1971

6.

7. Author(s)

8. Performing Organization Rept. No.

9. Performing Organization Name and Address
Consulting Division
Chemical Construction Corporation
320 Park Avenue
New York, New York 10022

43-A-5

10. Project/Task/Work Unit No.

11. Contract/Grant No.
CPA 22-69-151

12. Sponsoring Organization Name and Address
Division of Process Control Engineering
National Air Pollution Control Administration
Environmental Health Service
U. S. Department of Health, Education and Welfare
Washington, D. C. 20201

13. Type of Report & Period Covered

14.

15. Supplementary Notes DISCLAIMER- This report was furnished to the Office of Air Programs by Consulting Division, Chemical Construction Corporation, 320 Park Avenue, New York, New York 10022 in fulfillment of CPA 22-69-151

16. Abstracts

The pyritic sulfur content of some bituminous coals can be reduced by gravimetric separation or "washing". The extraction of energy and sulfur value contained in the reject fractions is technically feasible. Under favorable conditions the sale of extracted energy and sulfur products will to some degree offset the cost of washing. The purpose of this study is to select, design, and evaluate the high sulfur combustor and sulfur recovery process with the greatest potential for utilizing high-ash, high-sulfur coal reject material produced by the deep cleaning of coal. The report therefore includes preliminary designs and evaluations of several combustion and flue gas treating systems for processing high sulfur fuels drawn off from rejects of coal washing. These fuels, will need to be made up to specifications that depend on; the requirements of available processes for sulfur value recovery, and on the limitations of combustion equipment in respect of ash. The "HSC" Fuel specifications can be satisfied by appropiately re-combining selected portions of reject material from deep cleaning of coal. Four specimen HSC Fuel compositions are tabulated along with product and cost data on six processing systems for these fuels.

'High Sulfur Content '(HSC)

17. Key Words and Document Analysis. 17a. Descriptors

Air pollution
Desulfulfurization
Pyrite
Coal
Coal preparation
Washing
Cleaning
Market value
Materials recovery
Sulfur
Energy
Combustion chambers
Expenses
Furnaces
Electric power generation
Sulfur dioxide

17b. Identifiers/Open-Ended Terms
Air pollution control

17c. COSATI Field/Group 13B

18. Availability Statement Unlimited

19. Security Class (This Report) UNCLASSIFIED

20. Security Class (This Page UNCLASSIFIED

21. No. of Pages 223

22. Price

A

FORM NTIS-35 (10-70)

USCOMM-DC 40329-P71

C

APTD-0768

THE HIGH SULFUR COMBUSTOR

A Study of Systems
for
Coal Refuse Processing

VOLUME ONE
(NARRATIVE SUMMARY)

FINAL REPORT

to

DIVISION OF PROCESS CONTROL ENGINEERING
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION
ENVIRONMENTAL HEALTH SERVICE

U.S. DEPARTMENT
of
HEALTH, EDUCATION and WELFARE

CONTRACT NO. CPA 22-69-151

February 1971

Consulting Division
Chemical Construction Corporation
320 Park Avenue
New York, New York 10022

237

FOREWORD

Chemical Construction Corporation with assistance from Foster Wheeler Corporation, Monsanto Enviro-Chem Systems Inc., and Ebasco Services, Inc., has conducted a study of processing systems for extracting energy and sulfur values from the refuse of coal cleaning. This is the Final Report submitted in accordance with Contract No. CPA 22-69-151 dated June 30, 1969 as amended.

The Report is bound in two volumes, of which Volume One (Parts I through X) is essentially a summary for highlight purposes, and Volume Two (Parts XI through XVII) in which most of the study detail is compiled.

HIGH SULFUR COMBUSTOR STUDY

CONTENTS

VOLUME ONE - NARRATIVE SUMMARY

Page

PART I

ABSTRACT
INTRODUCTION I-1
1. CONTRACT OBJECTIVE
2. STATEMENT OF CONTRACT WORK I-2
3. CRITERIA FOR PROCESS SELECTION I-4
Table I-1 - Criteria for Process Classification I-5
Table I-2 - Factors for Ranking Sulfur Oxides Removal Processes
4. GEOGRAPHIC CRITERIA I-6
5. SITE CRITERIA I-7
6. PRINCIPAL BACKGROUND LITERATURE I-8
Figure I-1 - Pyrite-Coal Program I-9
7. ACKNOWLEDGMENTS AND DOCUMENTATION I-10

PART II

COAL PREPARATION II-1
1. PRESENT PRACTICE
2. THE OCCURRENCE OF SULFUR IN COAL II-2
3. COMMERCIAL TECHNIQUES II-4
4. DEEP CLEANING II-5
5. DEVELOPMENT ACTIVITY II-6
5.1 Current and Completed Studies
Figure II-1 - Pyritic Sulfur Reduction II-9
Figure II-2 - Total Sulfur Reduction II-10
6. FLEXIBILITY OF CLEANING METHODS II-11
Figure II-3 - Flow Sheet for Concentrating Table Tests II-13

Page

PART II

COAL PREPARATION (Continued)

Figure II-4 - Flow Sheet for Pyrite Beneficiation Tests II-14
Figure II-5 - Pyrite Beneficiation Circuit II-15
Tables II-1 through II-5 - Concentrating Table Test Results II-17 thru II-21
7. PROCESS CONTROL OF REJECT STREAMS II-22
    7.1 Tailor-made HSC Fuels
8. VALUE OF HSC FUELS

PART III

BASE PARAMETERS III-1
1. ENERGY EXTRACTION
    1.1 Sulfur/Carbon Ratio
    Figure III-1 - Combustion of Sulfur/Carbon Mixtures in 25% Excess Air III-2
    1.2 Heating Value of HSC Fuel III-3
    1.3 Limits of Study
    1.4 Three Process Systems Examined
    1.4.1 through 1.4.5 - Five Cases III-5 thru III-6
    Figure III-2 - Coal, Energy and Sulfur Products III-7
    1.5 Product Balance III-8
2. SULFUR VALUE III-9
    2.1 Sulfuric Acid
    2.2 Sulfur III-10
    2.3 Magnesium Sulfite and Liquid Sulfur Dioxide
    Figure III-3 - Consumption of Sulfur in the U.S. III-11
    2.4 Trade-Off Basis III-12

Page

PART IV

FIRING OF HSC FUELS IV-1

1. FUEL COMPOSITIONS

1.1 Whole Refuse Fraction

Table IV-1 - Whole Refuse Fraction from Five ROM Coals IV-2

1.2 Blending of Fractions IV-3

1.3 Heat Value Yield and Non-Pyritic Ash

1.4 Ash Fusion Temperature IV-4

1.5 Four HSC Fuel Specifications

Figure IV-1 - Ash Softening Temperature Depending on Basic Content of Coal Ash IV-5

Table IV-2 - HSC Fuel Composition CASE A IV-6

Table IV-3 - HSC Fuel Composition CASE B IV-7

Table IV-4 - HSC Fuel Composition CASE C and D IV-8

Table IV-5 - HSC Fuel Composition CASE E IV-9

2. FURNACE TYPES IV-10

2.1 Roasting Equipment

2.2 Fluidized Bed Combustors

2.3 Stoker Furnaces

2.4 Grate Furnaces

2.5 Crushed Coal-Fired

2.6 Pulverized-Fuel Fired

3. FURNACE LOAD REQUIREMENTS IV-11

3.1 Steam Design

3.2 Service Factor

4. FURNACE SELECTION BASIS

4.1 Dry Bottom Suspension Combustion IV-12

Page

PART IV (Continued)

FIRING OF HSC FUELS

5. COMBUSTION PROCESS DESIGN IV-13

Figure IV-2 - Combustion Process Flow Sheet IV-14

Table IV-6 - Expected HSC Fuel-Fired Combustor Performance

Figure IV-3 - Combustion Effects of Pyrite Mixtures

Table IV-7 - Summary of HSC Combustion Processes IV-16

6. INDUSTRIAL BOILER IV-17

6.1 Vertical Firing

6.2 Steam Characteristics

Figure IV-4 - 500,000 LB/HR Industrial Type High Sulfur Combustor, Sectional Side Elevation IV-18

Figure IV-5 - Foster Wheeler Anthracite Burner IV-19

6.3 Energy Distribution IV-20

Figure IV-6 - Distribution of Energy in 500,000 LB/HR Boiler for CASE A IV-21

7. THE 50 MEGAWATT PROTOTYPE BOILER FOR CASE C IV-22

7.1 Horizontal Firing

7.2 Steam Characteristics

Figure IV-7 - 50 MW Prototype High Sulfur Combustor, Sectional Side Elevation IV-23

Figure IV-8 - Horizontal Intervane Burner Equipped for Pulverized Coal, Gas and Oil Firing IV-24

7.3 Energy Distribution IV-25

Figure IV-9 - Distribution of Energy in the 50 MW Boiler for CASE C IV-26

Page

PART IV (Continued)

FIRING OF HSC FUELS

8. THE 500 MEGAWATT ELECTRIC UTILITY BOILER IV-27
 8.1 Horizontally Fired
 8.2 Steam Characteristics
 Figure IV-10 - The 500 MW High Sulfur Combustor Sectional Side Elevation IV-28
 8.3 Energy Distribution IV-29
 Figure IV-11 - Distribution of Energy in the 500 MW Boiler for CASE D IV-30
9. BOTTOM ASH AND FLY ASH IV-31
 9.1 Iron in Ash
 Table IV-8 - Fly-Ash Collection Design Basis IV-32

PART V

ELECTRIC GENERATION

1. SIZING OF STEAM-ELECTRIC EQUIPMENT V-1
2. CONVENTIONAL TURBINE CYCLE V-2
3. CENTRAL STATION STEAM GENERATORS
 Figure V-1 - Pulverized HSC Fuel-Fired Steam/Power Plant V-3
 Table V-1 - List of Central Station Steam Generators V-4 thru V-17

PART VI

SULFUR VALUE RECOVERY VI-1

1. FLUE GAS CLEANING
2. MARKETABLE END-PRODUCTS
3. STRONG CHEMISTRY WANTED VI-2
 3.1 High Sulfur Value Yield
 3.2 Adaptability
 3.3 Minimum Pollution Risk

Page

PART VI (Continued)

SULFUR VALUE RECOVERY

4. AVAILABLE PROCESSES

4.1 The "Contact" Process VI-3
  4.1.1 Standard Process
  4.1.2 The Cat-Ox Process VI-4
4.2 Other Sulfuric Acid Processes VI-5
4.3 Concentration of Weak Gases to Intermediate Levels
  4.3.1 Reinluft Process VI-6
  4.3.2 Grillo Process
  4.3.3 Magnesium Sulfite/Oxide Process VI-7
4.4 Concentration of Weak Gases to High (90/100%) Level VI-9
  4.4.1 Aromatic Amine Absorbents VI-10
  4.4.2 Basic Aluminum Sulfate VI-11
  4.4.3 Thermal Cycle With Water VI-12
  4.4.4 Alkali-Metal Sulfite/Bisulfite VI-13
4.5 Liquefaction of SO_2 VI-14
4.6 Processes for Output of Elementary Sulfur
  4.6.1 Reduction of 100% SO_2 by Hot Coke VI-15
  4.6.2 Reduction of Dilute SO_2 with Producer Gas VI-16
  4.6.3 Reduction of SO_2 with Methane VI-17
    4.6.3.1 Non-catalytic processes
    4.6.3.2 Catalytic processes VI-18
  4.6.4 Reduction of Dilute SO_2 with Reformed Natural Gas VI-19
  4.6.5 Reduction of SO_2 with H_2S VI-20

Page

PART VI (Continued)

SULFUR VALUE RECOVERY
5. EMISSION CONTROL
5.1 Control of SO_2 in Tail Gas VI-22
5.2 Control of Ambient SO_2
5.3 Control of Particulates

PART VII

SUMMARY OF HSC SYSTEMS VII-1
1. SUMMARY OF THE CASE A SYSTEM
Figure VII-1 - Summary Flow Sheet CASE A VII-2
2. SUMMARY OF THE CASE B SYSTEM VII-3
Figure VII-2 - Summary Flow Sheet CASE B VII-4
3. SUMMARY OF THE CASE C PROTOTYPE SYSTEM VII-5
Figure VII-3 - Summary Flow Sheet CASE C VII-6
4. SUMMARY OF THE CASE D SYSTEM VII-7
Figure VII-4 - Summary Flow Sheet CASE D VII-8
5. SUMMARY OF THE CASE E SYSTEM VII-9
Figure VII-5 - Summary Flow Sheet CASE E VII-10

PART VIII

ECONOMIC EVALUATION
1. METHOD
1.1 Valuation Formula VIII-1
1.2 Precombustion Costs vs. Dumping VIII-2

Page

PART VIII (Continued)

ECONOMIC EVALUATION

2. COSTS, PRICES AND ESCALATION VIII-3
  2.1 Capital Costs
  Table VIII-1 - Estimated Capital Cost
  2.2 Costs of Operation or Extraction VIII-4
    2.2.1 Utilities
    2.2.2 Operating Labor and Supervision
    2.2.3 Maintenance Cost VIII-5
    2.2.4 Local Taxes, Insurance and the Cost of Money
    2.2.5 Depreciation VIII-5
    2.2.6 Product Costing Basis VIII-6
    Tables VIII-2 through VIII-6 - Estimated Operating Costs of HSC Systems VIII-7 thru VIII-11
    2.2.7 Pollution Control VIII-12
    Table VIII-7 - Pollution Control as Alternate to Sulfur Value Recovery VIII-13
  2.3 Sales Value of Products VIII-14
    2.3.1 Steam
    2.3.2 Electricity VIII-15
    2.3.3 Sulfur
    Table VIII-8 - Sulfuric Acid - Estimated Production and Shipments 1968 VIII-19
    2.3.4 Sulfuric Acid VIII-20
      2.3.4.1 Major end-uses
      2.3.4.2 Sulfuric acid list prices VIII-24
      Table VIII-8 - Sulfuric Acid - Average price of Shipments
      Table VIII-9 - Sulfuric Acid uses in the United States VIII-27

Page

PART VIII (Continued)

ECONOMIC EVALUATION

3. OPERATING ECONOMICS VIII-28

Table VIII-10 - Income from Sales, Costs of Operation and Products, Gain or Loss from Operation

4. VALUE OF HSC FUELS AND PAYOUT VIII-30

Table VIII-11 - Value of HSC Fuels, Offset to Cleaning Cost, Payout of Investment VIII-31

5. SHIPPING ECONOMICS VIII-32

5.1 End Product: Sulfuric Acid Delivered

5.2 Analysis of Sulfuric Acid Delivered Cost VIII-33

Figure VIII-1 - Linear Model for Analysis of Production and Shipping Economics

5.2.1 Input data: Production, Storage and Transportation Costs VIII-34

Table VIII-12 - Tabulation of Input Data

5.2.2 Solution favors CASE D3 and barge VIII-36

Table VIII-13 - Printout of Analysis

Table VIII-14 - Optimal Choice Index with Barge VIII-39

Table VIII-15 - Optimal Choice Index without Barge VIII-41

6. CASE D VS. CASE D3 VIII-42

Table VIII-16 - D and D3 Capital and Operating Costs and Gain Compared VIII-43

Page

PART IX

ABSTRACTS OF REFERENCE LITERATURE 1 to 23

PART X

SUBJECT INDEX TO REFERENCE LITERATURE

HIGH SULFUR COMBUSTOR STUDY

CONTENTS

VOLUME TWO - DESCRIPTIVE DETAIL

(Separately bound)

PART XI

FUEL SPECIFICATION AND COMBUSTOR DESIGN
(Foster Wheeler Corporation)

PART XII

ASH

PART XIII

CASE A SULFURIC ACID
- Process Description
- Process Flow Sheet
- Summary of Capital and Operating Cost

PART XIV

CASE B SULFURIC ACID
- Process Description
- Process Flow Sheet
- Summary of Capital and Operating Cost

PART XV

CASE C PROTOTPYE SULFUR
- Process Description
- Process Flow Sheet
- Equipment List and Specification
- Capital Cost Estimate
- Summary of Capital and Operating Cost

PART XVI

CASE D SULFUR

Process Description
Process Flow Sheet
Summary of Capital and Operating Cost

PART XVII

CASE E SULFURIC ACID

Process Description
Process Flow Sheet
Summary of Capital and Operating Cost

High Sulfur Combustor Study

ABSTRACT

The pyritic sulfur content of some bituminous coals can be reduced by gravimetric separation or "washing". The extraction of energy and sulfur value contained in the reject fractions is technically feasible. Under favorable conditions the sale of extracted energy and sulfur products will to some degree offset the cost of washing. The report includes preliminary designs and evaluations of several combustion and flue gas treating systems for processing high sulfur fuels that may be drawn off from the rejects of coal washing. These fuels will need to be made up to specifications that depend first on the requirements of available processes for sulfur value recovery, and second on the limitations of combustion equipment in respect of ash. The "HSC" Fuel specifications can evidently be satisfied by appropriately re-combining selected portions of reject material from deep cleaning of coal. Four specimen HSC Fuel compositions are tabulated on the following page together with product and cost data on six processing systems for these fuels. The operating context in each case is one of a number of typical site possibilities in the bituminous coal-producing area comprised of Pennsylvania, Ohio, West Virginia, Kentucky, Indiana and Illinois.

HSC FUEL COMPOSITIONS, PRODUCTS, INVESTMENT AND OPERATING COSTS

	CASE A	CASE B	CASE C PROTOTYPE	CASE D	CASE D3	CASE E
High Sulfur Fuel Composition						
% Total Sulfur	33.1	18.3	11.7	11.7	11.7	7.9
% Total Ash	61.0	42.5	33.8	33.8	33.8	28.8
Sulfur to Coal Weight Ratio	1.8	0.4	0.2	0.2	0.2	0.12
Percent SO_2 in Flue Gas	6.0	2.0	1.2	1.2	1.2	0.7
Products for Sale						
Energy as Steam - M Lbs/Hr	500	500	500	3,500	3,500	3,500
Energy as Electric Capacity - Megawatts			50	500	500	500
Sulfur Value as Sulfuric Acid - M Net Tons per Year	610	208				
Sulfur Value as Sulfur - M Long Tons per Year			31	233	753	473
Co-Produced Clean Coal M Tons per Year	7,100	2,300	920	7,100	7,100	3,500
TOTAL INVESTMENT IN HSC SYSTEMS	M$18,340	M$18,600	M$29,270	M$128,520	M$126,520	M$125,700
ANNUAL COST OF OPERATION BEFORE U.S. INCOME TAX	6,990	6,060	7,530	34,500	29,000	27,900
COST OF ENERGY EXTRACTION						
Steam per M Lbs	$ 3.00	$ 1.60				
Electricity per MWH			$ 14.50	$ 7.00	$ 7.00	$ 5.80
COST OF SULFUR VALUE EXTRACTION						
Sulfuric Acid per Net Ton	($ 8.20)	($ 2.00)				
Sulfur per Long Ton			$ 57.00	$ 28.00	$ 3.50	$ 10.00

Note: "Cost" does not include any charge for the HSC Fuels. Unit costs of products reflect an adjustment for "pollution control". See PART VIII 2.2.7.

In CASE A and CASE B the cost of sulfuric acid manufacture is less than the cost of pollution control, hence the figures in ( ) parenthesis. See TABLE VIII-10.

Forward estimates of selling prices of the four products are:

Steam per M lbs	$ 0.65
Electricity per KWH	0.0075
Sulfur per LT	25.00
Sulfuric Acid per NT	12.75

If these prices prevail, the estimates of economic benefit, if any, treated as credits to coal cleaning costs, are as shown below:

CASE	A M$	B M$	C M$	D M$	D3 M$	E M$
Gain after Tax	1,760	(420)	(1,960)	690	4,700	4,230
Clean Coal: M Ton/Year	7,100	2,300	920	7,100	7,100	3,500
Netback per Ton of Clean Coal	$0.25	($0.18)	($2.13)	$0.10	$0.66	$1.21

CASE C is intended for possible construction as a prototype or demonstration plant. The four other case studies, taken together, amount to a preliminary exploration of the range of possible operating conditions. Any specific coal cleaning situation will need to be examined in detail for applicability of the high sulfur combustor idea. Washability of the coal and proximity to sulfuric acid consumers are shown to be major determinants of the choice of technology and economic expectations. The manufacture of sulfur for conversion after shipment to sulfuric acid compares unfavorably with manufacture and shipment directly of sulfuric acid unless very large tonnages are to be shipped long distances by rail.

THE HIGH SULFUR COMBUSTOR

I

INTRODUCTION

1. "The Division of Process Control Engineering of the National Air Pollution Control Administration is developing processes for the control of air pollution from stationary sources. A promising technique for the control of sulfur dioxide pollution from coal-burning installations is the removal of pyritic sulfur from the coal prior to combustion. Although technology is available for "deep" cleaning of coal for pyrite removal, the economics of deep cleaning has prevented widespread use of the technology for cleaning so-called "steam coals". Significant amounts of carbon are lost in the reject material from the coal-cleaning plant. A potential solution to the problem of carbon loss is the utilization of this high-ash, high-sulfur, carbon-bearing reject material in a special combustor. The reject material should have sufficient heat value for steam raising, and the flue gas resulting from the combustion of the reject material should contain a relatively high concentration of sulfur compounds. Recovery of these compounds by the flue gas processing would provide a useful and salable product. Feasibility studies conducted previously by NAPCA have indicated that a high sulfur content in the flue gas may provide the economic incentive for deep cleaning of steam coals. A process that would produce sulfuric acid is only one of the processes with potential value; another promising approach would produce elemental sulfur.

"Deep coal cleaning combined with utilization of reject materials in a high-sulfur combustor with associated sulfur recovery system needs to be evaluated. NAPCA has already underway a program in which a prototype 500- to 1,000-tpd coal-cleaning plant will be built and a wide range of coals subjected to cleaning. The economics and technical feasibility of the coal-cleaning process will thus be established. Selection, design, and evaluation of a high-sulfur combustor and by-product recovery system needs to be accomplished to prove the economic feasibility of deep cleaning of steam coals.

"The purpose of this study is to select, design, and evaluate the high-sulfur combustor and sulfur recovery process with the greatest potential for utilizing high-ash, high-sulfur coal reject material produced by the deep cleaning of coal. The economic and technical feasibility of such a process will be established by this study."

2. STATEMENT OF CONTRACT WORK

2.1 **"Survey, analyze, select, and design a combustor system that will utilize high-ash, high-sulfur, coal reject material produced by the deep cleaning of coal as fuel for the production of steam for power generation or process use.**

2.2 **"Select, analyze, and design a flue gas treatment process that will recover sulfuric acid or sulfur from the SO_2-rich off-gases from the combustor. Primary consideration should be given to commercial processes or those**

that are nearly commercial. In addition, recognition should be given to marketability of sulfur products.

2.3 "Determine those chemical and physical properties of the reject materials that will influence the design of the combustor system.

2.4 "Establish the tests and procedures required to characterize the reject material for use as a fuel.

2.5 "Coordinate the efforts of the current NAPCA coal-cleaning program with this proposed study. Specifically, correlate with the coal-cleaning program on those specifications of the cleaning reject material that will make the material usable as a fuel, and determine appropriate composition ranges that are required and can be met by coal preparation techniques. In addition, recommend for the coal-cleaning plant additional cleaning circuits that will result in the production of a reject material of desired composition.

2.6 'Perform a conceptual design of a prototype plant and cost estimate of the overall process, including coal cleaning, the high-sulfur combustor for power/steam generation, and the flue gas sulfur recovery system. Specifications range of the reject material for use as a fuel will also be established. The results of this task will provide the basis for decision whether to build a high-sulfur combustor and sulfur recovery prototype plant. Size of prototype plant will be such as to demonstrate commercial applicability.

2.7 "A final task will be the extrapolation of data to conceptual designs with cost estimates for full-scale industrial and utility boilers (combustors), such as, a 500,000 pounds per hour (steam) industrial boiler and a 500-megawatt utility station, both with sulfur recovery systems."

3. CRITERIA FOR PROCESS SELECTION

Due attention has been paid to the emphasis in Item 2.2 on the selection of "commercial or nearly commercial processes" to serve as many functions as possible of the combustor and SO_2 recovery systems under study. Such processes are considered to be in the front rank of availability if otherwise suitable for these systems. Where commercial processes have appeared to be inaccessible or nonexistent for some functions of the systems, recourse has been had to processes not yet commercialized. The selection of non-commercial processes for analysis in this study has been guided by the conception of diminishing "availability" in Table I-1 and the "ranking factors" of Table I-2.

TABLE I-1

"CRITERIA FOR PROCESS CLASSIFICATION

A. FIRST GENERATION

(1) Advanced pilot plant studies
(2) Active research and development
(3) Adaptability to U.S. market
(4) Available data on economic assessment

B. NEAR FIRST GENERATION

(1) General first generation criteria but:
(2) Less advanced pilot studies
(3) More interest and better adaptation by foreign nations
(4) Sulfur by-products not attractive to U.S. market

C. SECOND GENERATION

(1) Essentially bench studies
(2) Minimal pilot plant operations
(3) New concepts represented
(4) Process potential may exceed that of first generation methods
(5) Economic assessment not yet practical"

(Ref. EE)

TABLE I-2

"FACTORS FOR RANKING SULFUR OXIDES REMOVAL PROCESSES

TECHNICAL FEASIBILITY
    STATE OF DEVELOPMENT
        PROCESS APPLICABILITY
            RELATIVE ADVANTAGES
                PROBLEM AREAS
                    ECONOMIC FACTORS"

(Ref. EE)

3.1 The clean coal and the energy and sulfur values extracted from the reject material should be products of commercial grade and realized in well balanced quantities so that none of the three is outsize in terms of the market expected to absorb it, or of too small a tonnage for good production economy.

3.2 Combustor design for steam-electric generation should be conservative and for base-loading.

3.3 Combustor design for 500,000 pounds per hour of steam output should be of industrial type with good turndown characteristics.

3.4 Processes for recovery of sulfur value should be capable of high turndown ratios.

4. GEOGRAPHIC CRITERIA

Where necessary in this study for argument purposes to visualize a high sulfur combustor system operating in a geographic context, the frame of reference is the six-state area of Pennsylvania, Ohio, West Virginia, Kentucky, Indiana and Illinois. These states are the principal producers of steam coal for electric utilities; they are also consumers of electric energy in abundance, and moderate consumers of sulphuric acid. The city of Cincinnati, roughly central to this area, has served the study as a point of reference for certain shipping costs and air pollution criteria.

5. At such a time as requires selection of an actual site on which to locate a project of a type discussed in this study, consideration may be given to:

   5.1 The site of the mine expected to supply ROM coal. Such a site might be preferred if, for example, coal cleaning for sulfur reduction requires moderate crushing only so that the low-sulfur clean coal is not too fine for shipping.

   5.2 The site of an existing electric utility complex to which raw coal would be shipped and in which the cleaned coal would be consumed.

   5.3 The site of a large chemical works, consumer of steam and sulfuric acid, to which the special high sulfur fuel would be shipped.

   5.4 A site on navigable water since water shipment of any or all of the bulk coal and sulfur products will usually yield an advantage.

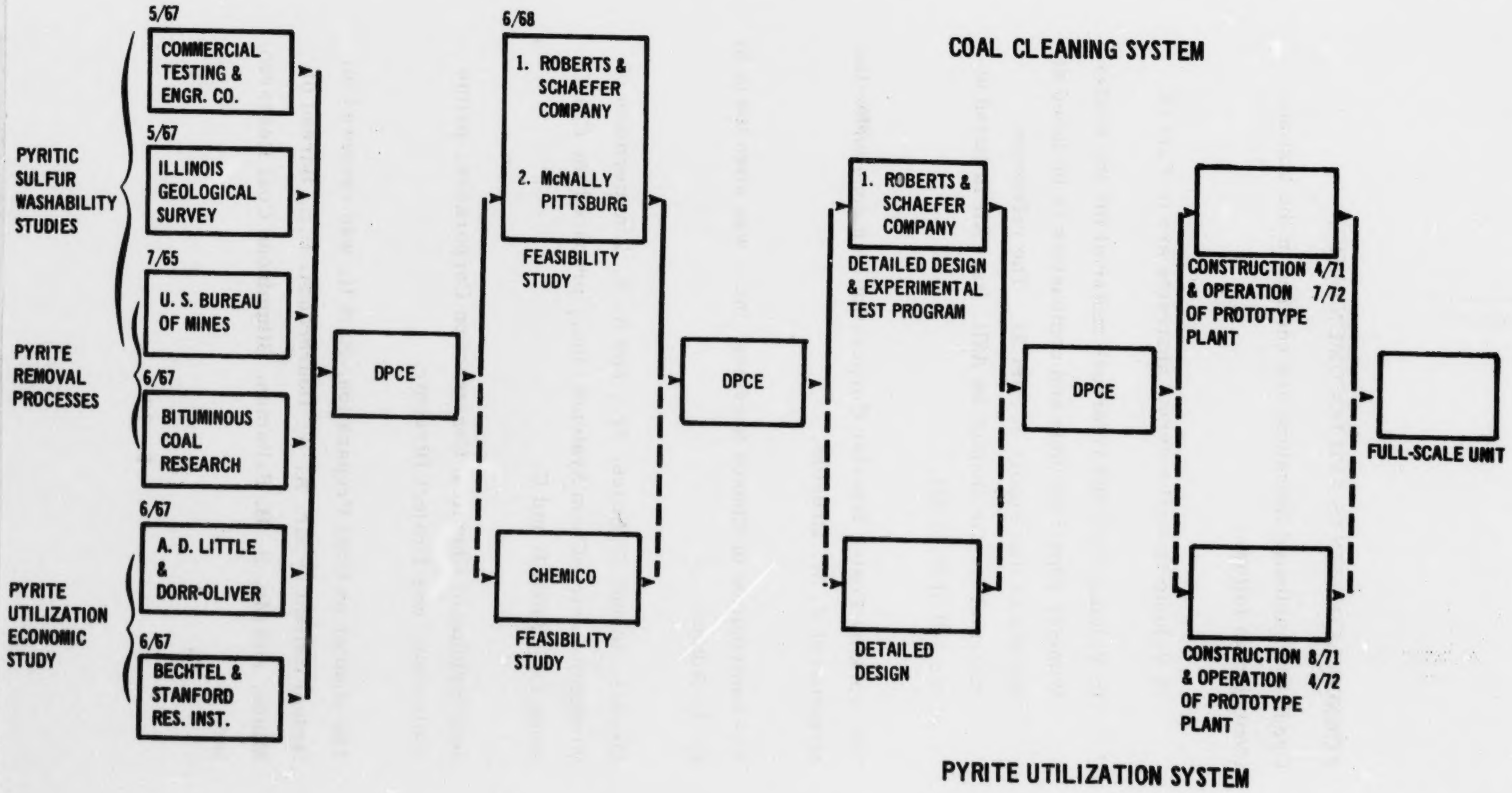
It is especially to be noted that the coal processing systems discussed in this report are all dependent on some degree of initial crushing and grinding of the ROM. Since shipping of finely ground coal by rail in open gondolas is costly as to loss by windage, the choice of operating site may be consequently much influenced. There is considerable evidence from current experimental work that milling of coal to "pulverized coal" size prior to washing will in most cases permit a greater sulfur reduction than can be effected

on less finely milled coal. Where this is true of a particular coal to be processed, a site in the first or second of the categories above would seem to gain an advantage.

6. PRINCIPAL BACKGROUND LITERATURE

Several recent investigations in the coal cleaning arts, directed particularly at reduction of pyritic sulfur and processing of the high sulfur refuse, now underway or completed by the Bureau of Mines, by others, and by private organizations under contract to NAPCA, together with a number of subventions by NAPCA of university research, represent a very substantial body of work accomplished in support of continuing studies in this field. The Program is illustrated in Figure I-1 following this page.

FIGURE I-1


COAL CLEANING SYSTEM
PYRITIC SULFUR WASHABILITY STUDIES
PYRITE REMOVAL PROCESSES
PYRITE UTILIZATION ECONOMIC STUDY
5/67
COMMERCIAL TESTING & ENGR. CO.
5/67
ILLINOIS GEOLOGICAL SURVEY
7/65
U. S. BUREAU OF MINES
6/67
BITUMINOUS COAL RESEARCH
6/67
A. D. LITTLE & DORR-OLIVER
6/67
BECHTEL & STANFORD RES. INST.
DPCE
6/68
1. ROBERTS & SCHAEFER COMPANY
2. McNALLY PITTSBURG
FEASIBILITY STUDY
CHEMICO
FEASIBILITY STUDY
DPCE
1. ROBERTS & SCHAEFER COMPANY
DETAILED DESIGN & EXPERIMENTAL TEST PROGRAM
DETAILED DESIGN
DPCE
CONSTRUCTION 4/71
& OPERATION 7/72
OF PROTOTYPE PLANT
CONSTRUCTION 8/71
& OPERATION 4/72
OF PROTOTYPE PLANT
FULL-SCALE UNIT
PYRITE UTILIZATION SYSTEM

NAPCA-DPCE* PYRITE COAL PROGRAM

7. ACKNOWLEDGMENTS AND DOCUMENTATION

Credits to published literature are entered in the text and identified as follows:

> In Volume One all referent abstracts are in Part IX.
>
> In Volume Two, the reference material for the Foster Wheeler report on fuels and combustors is included at the end of that report in Part XI. The reference material for the chapter on ASH, Part XII is listed at the end of Part XII.

The work by Foster Wheeler Corporation was done under the direction of R. W. Breyers.

The contribution of Ebasco Services, Inc., was attended to by P. J. Adams.

Messrs. Joseph G. Stites, Jr., and R. E. Zimmermann of Monsanto Enviro-Chem Systems, Inc., prepared the Cat-Ox study for CASES B and E.

John Belding of Chemical Construction Corporation, prime contractor, was Project Director.

The chapter on Coal Preparation, Part II, was reviewed for factual content by Mr. A. W. Deurbrouck, U.S. Bureau of Mines, and Mr. R. B. Saltsman, Bituminous Coal Research, Inc.

The National Air Pollution Control Administration was represented by Mr. T. Kelly Janes and Mr. G.S. Haselberger, Division of Process Control Engineering, who served the project successively as Project Officers. Mr. R.P. Hangebrauck of this Division was also especially interested and helpful.

Contributions of value have been made by many people not mentioned by name.

II

COAL PREPARATION

This term is applicable to any routine modification of a mined coal that is intended to improve acceptance by the market or improve suitability of the coal for a specific use. For the present purpose interest is centered on modifications of coal that effect reduction of sulfur content but questions of heat value yield, ash characteristics, sizing, etc. are, of course, also involved.

1. PRESENT PRACTICE

Coal now supplied to utilities is variously processed depending on ROM characteristics and customers' requirements. Emphasis on high heating value consistent with yield has followed increased costs of shipping and stiffer competition from other fuels. Coal "washing" to upgrade heat value has been practiced many years, and to some degree elimination of the pyritic source of ash (ferric oxide) has figured in this practice over a long time.

But "Generally, coal preparation plants for cleaning steam coals are not designed for maximum desulfurization. The much smaller quantity of coal for metallurgical coke requires lower sulfur limits. As a result, some of the steel companies have developed some of the most efficient plants for desulfurization", Ref. (B). Steam coal producers certainly have access to most of the technology developed for ore dressing and metallurgical coal beneficiation, but applications to steam coal have been limited by lack of economic incentive.

2. THE OCCURRENCE OF SULFUR IN COAL

"Sulfur occurs in coal in three principal forms: sulfate, organic, and pyritic. The amount of sulfate sulfur in freshly mined coal is normally small and of little significance. The organic sulfur in coal is more or less uniformly distributed throughout the coal substance in molecular combination, and cannot be removed even partially without materially altering the nature of the coal substance. The organic sulfur content of coal ranges from a low of 20 per cent to a high of 60 per cent* of the total sulfur. Obviously, the organic sulfur content of a coal is the prime parameter in determining whether or not significant total sulfur reductions can be made with a given coal by pyrite removal.

"Pyrite occurs in coal as discrete particles in a wide variety of shapes and sizes.

"The principal forms are:

2.1 Rounded masses called "sulfur balls" or nodules which range in size from a small fraction of an inch to very large.

2.2 Lens-shaped masses which may be thought of as flattened sulfur balls which vary greatly in thickness and lateral extent.

2.3 Veins of pyrite consisting of vertical or inclined veins or fissures filled with pyrite ranging in thickness from thin flakes up to several inches thick in some cases.

---

* Other observors report a high of 80 per cent.

2.4 **Small discontinuous veinlets of pyrite**, a number of which sometimes radiate from a common center which may be a small sulfur ball.

2.5 **Small particles or veinlets** disseminated in the coal.[+]

"All coals contain forms .3 and .5, and some coals contain all five of the principal forms.

"The total amount of pyritic sulfur varies greatly in different coal beds, and also in different mines operating in the same coal bed.

"Likewise, there is great variation in the size and shape of pyrite inclusions in coals having a similar amount of total pyritic sulfur. The degree of pyrite liberation at any given stage of crushing and grinding also varies greatly among coals. With some coals, good pyrite removal is obtained by merely washing the nominal sizes produced in the mining operations, and little additional pyrite is liberated by further size reductions. Other coals show some pyrite liberation with each successive stage of reduction, while still other coals show no significant pyrite liberation when they are pulverized to conventional p.c.[*] size (60 to 85 per cent minus 200 mesh).

[+] These fine particles range from a size easily viewed with the naked eye to particles so small they they cannot be removed without materially altering the nature of the coal substance.

[*] Pulverized coal.

"These variations are directly related to the mode of occurrence of the pyrite; this is the second important parameter in determining whether or not sulfur reductions are feasible through pyrite removal. Coals which show little pyrite liberation when crushed to their p.c. size are not amenable to sulfur reduction through pyrite removal."

Ref. (KK) See also Ref. (N)

3. Some of the presently commercial coal processing techniques are as follows:

3.1 "Modified Run of Mine Coal

a) Large slate and large sulfur balls removed by hand picking.

b) Large slate and large sulfur balls or large fragments thereof removed during the size-reduction process utilizing a Bradford Breaker.

3.2 "Full Cleaning

a) Low cost wet cleaning of coarse sizes with a Baum Jig and low cost dry cleaning of smaller sizes with air tables.

b) More expensive but more efficient wet cleaning of coarse sizes with heavy media separators and wet fine coal cleaning on concentrating tables.

3.3 "Partial Cleaning

a) Coarse or fine coal cleaning with cleaned fraction being combined with the remainder in its run-of-mine condition." Ref (KK)

4. DEEP CLEANING

All of the processes mentioned will remove some of the liberated pyritic sulfur, but it is a substantial removal of sulfur from the cleaned coal that is needed for air pollution abatement, and a high concentration of sulfur in the reject fractions that best serves the combustor and recovery systems here visualized. This point of view is implicit in the newly coined term "deep cleaning" (with a connotation similar to "Full Cleaning" as in Item 3.2 above) by which, however, it is intended to identify processes directed at reduction of sulfur particularly. While deep cleaning is a rather new art, many of the techniques developed for beneficiation of other minerals, as mentioned earlier, can be adapted to this purpose. In addition to the float or sink separations in heavy media vessels of several designs, and filmsizing on wet concentrating tables, the techniques include concentrators such as the Humphrey spiral, water cyclones, air classifiers, etc. These and other devices, employed in laboratory development studies, effect separations of crushed ROM fractions by exploiting specific gravity differences which are, for example:

	Specific Gravity
Pyrite (FeS_2)	5.00
Marcasite (FeS_2)	4.87
Slate	2.5 to 2.8
Bituminous Coal	1.2 to 1.5

Froth flotation of coal fractions which does not depend on specific gravity, is also a promising technique for certain purposes.

5. DEVELOPMENT ACTIVITY

In recent years the Division of Process Control Engineering of NAPCA has sponsored a major program of coal washability studies directed at determining the supply of low sulfur coal recoverable by deep cleaning. "Pyrite washability studies have indicated significant potential for this air pollution control method. Of coals from 250 mines tested to date, 35 percent have organic sulfur levels of 1.0 percent or less and about half of the 35 percent could easily be cleaned to 1 percent or less total sulfur. An additional 25 percent of the coals have organic sulfur levels of 1.5 percent or less and of these about two-thirds could readily be cleaned to 1.5 percent total sulfur .....

5.1 "A list of studies completed or underway to determine the availability of washable coals and to develop a coal-cleaning and pyrite utilization process follows:

Feasibility Study of the Recovery of Sulfur and Iron From Coal Pyrites PB 176-844 Ref. (C)

An Economic Feasibility Study of Coal Desulfurization (Contractor: Paul Weir Company)
Vol. I, PB 176-845; Vol. II, PB 176-846 Ref. (E)

Occurrence and Removal of Pyritic Sulfur from American Coals
(U.S. Bureau of Mines)

Removal of Pyritic Coal in a Humphrey Spiral of Conventional and Modified Design
(U.S. Bureau of Mines)

Removal of Pyrite from Coal by Tabling
(U.S. Bureau of Mines)

Field Studies of Coal Cleaning Equipment
(U.S. Bureau of Mines)

Desulfurization of Coal by Froth Flotation
(U.S. Bureau of Mines)

Detailed Coal Analyses
(U.S. Bureau of Mines)

Electrokinetic Separation of Pyrite from Coal
(U.S. Bureau of Mines)

Characteristics and Removal of Pyritic Sulfur from American Coals
(U.S. Bureau of Mines)

An Evaluation of Coal Cleaning Methods and Techniques for Removal of Pyritic Sulfur from Coal
(Contractor: Bituminous Coal Research)
PH-86-67-139 Ref. (JJ)

Sampling and Evaluation of Coal Mines in Illinois
(Contractor: Illinois Geological Survey)
See also Ref. (T)

Evaluation of Float-Sink Testing of Raw Coal Samples from 20 Mines Selected by the NAPCA
(Contractor: Commercial Testing and Engineering Company)

Sampling and Testing of Coals from Selected Mines
(Contractor: Commercial Testing and Engineering Company)

Process Cost and Economics of Pyrite-Coal Utilization
(Contractor: Bechtel Corporation)
Ref. (B)

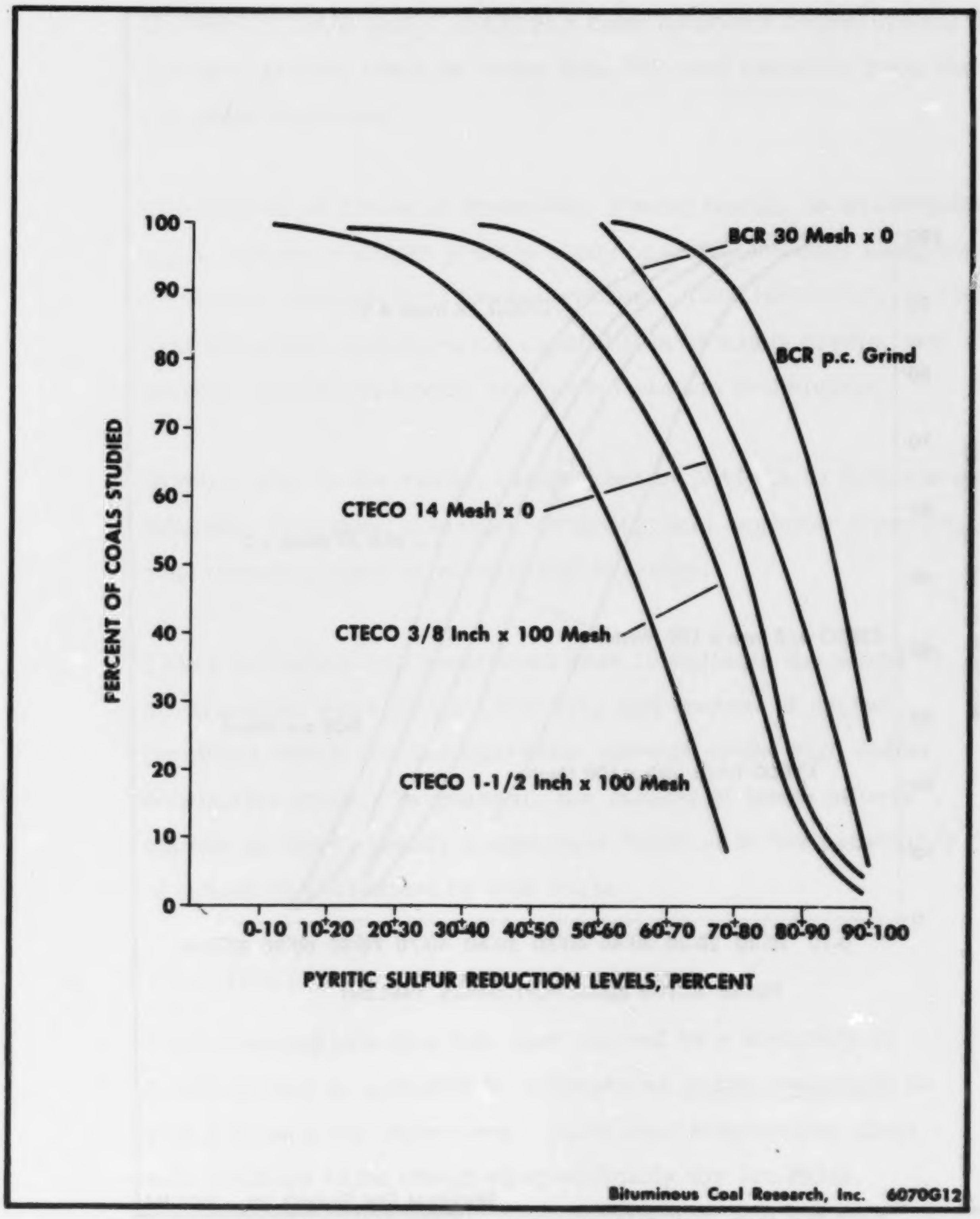
Study of Process Costs and Economics of Pyrite-Coal Utilization
(Contractor: A. D. Little, Inc.)
PB 182-303 Ref. (F)

Design and Cost Analysis Study for a Prototype Coal Cleaning Plant
(Contractors: McNally Pittsburgh Manufacturing Corporation
and
Roberts and Schaefer Company)"

Among the contractors listed above, Bituminous Coal Research, Inc. and Commercial Testing and Engineering Company collaborated on a study of samples of 70 coal reserves that are located entirely in the six-state area identified in Part I. The samples were tested by float-sink at 1.60 specific gravity as "related to nominal topsize of coal" for reduction of pyritic sulfur and reduction of total sulfur.

The results of this test program are plotted in Figure II-1 and Figure II-2.

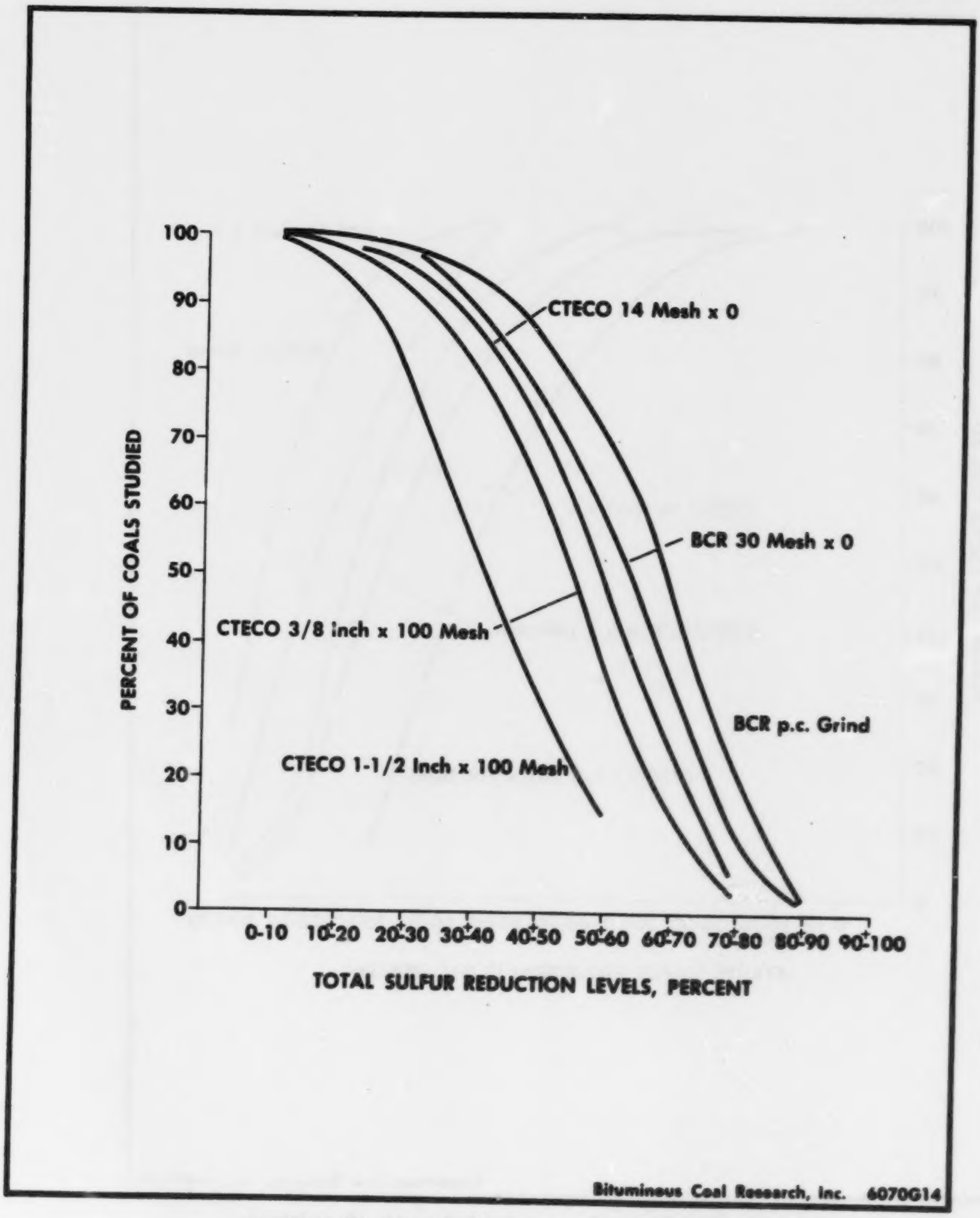
FIGURE II-1



Pyritic Sulfur Reduction at 1.60 Specific Gravity as Related to Nominal Topsize of Coal

Ref. (JJ)

FIGURE II - 2



Total Sulfur Reduction at 1.60 Specific Gravity as Related to Nominal Topsize of Coal

Ref. (JJ)

In addition to the foregoing test program at 1.60 specific gravity, the work of BCR and CTECO has been extended cooperatively to multiple gravity tests on more than 100 coal samples from the six-state reserves.

The Bureau of Mines at Bruceton, Pennsylvania, is accomplishing a similar multiple gravity study of some 200 coal samples to determine washability characteristics. This laboratory is also evaluating the desulfurizing capabilities of heavy media, wet tabling, spiral separator and froth flotation techniques.

Notable also is the recent assignment by NAPCA to Roberts and Schaefer Company, Chicago, to design and engineer a prototype coal cleaning plant of substantial capacity.

These activities are mentioned here to indicate the scope of development work on coal cleaning and control of reject fractions which are the upstream aspects of the high sulfur combustor study. In general, the results of these efforts appear so far to justify a continued interest in the possibility of pollution abatement by this route.

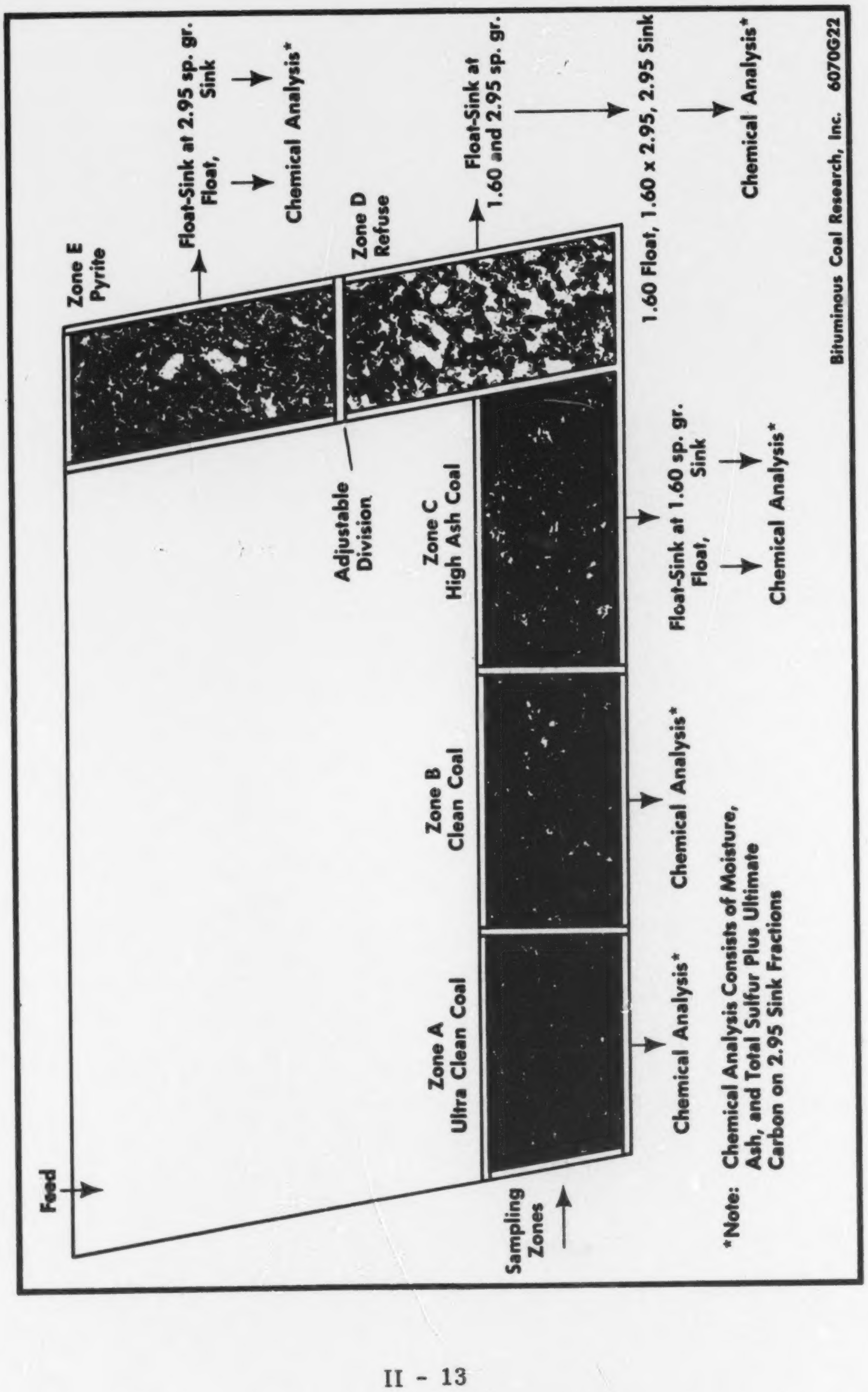
6. FLEXIBILITY OF CLEANING METHODS

Coal cleaning practice has been served by a diversity of methods that is expected to increase as sulfur reduction is added to existing objectives. Each coal preparation plant will continue to be designed specifically for the ROM feedstock available to it. Internal circuits will become more complex in the effort to maintain high yields, high heating value and maximum rejection of sulfur.

For example, the wet concentrating table appears to be typically versatile equipment. The table is a rhomboid plane surface inclined a few degrees and riffled at right angles to the incline. Crushed coal is introduced in a water slurry at an upper corner while the table is mechanically vibrated to stimulate travel of the coal. Low density particles tend to move directly down the incline, high density particles tend to move parallel to the riffles. Particles of intermediate density travel downward at corresponding intermediate angles. The charge leaves the table at the two edges opposite to the point of feed, see Figure II-3. The particles are classified as to increasing order of density all the way from lower left to upper right of the edges. Selected portions of the streams leaving the table can be diverted directly to shipment or to further processing by any chosen method.

As earlier noted, the wet table is only one of the many devices available to coal cleaners, but it well illustrates the potential of flexibility in separation techniques based on specific gravity differences. See, for example, Figure II-4, an experimental arrangement of three tables in series for pyrite concentration, or Figure II-5 in which concentrating tables, size classifiers and size reduction mills are alternated in the flow sheet.

FIGURE II-3

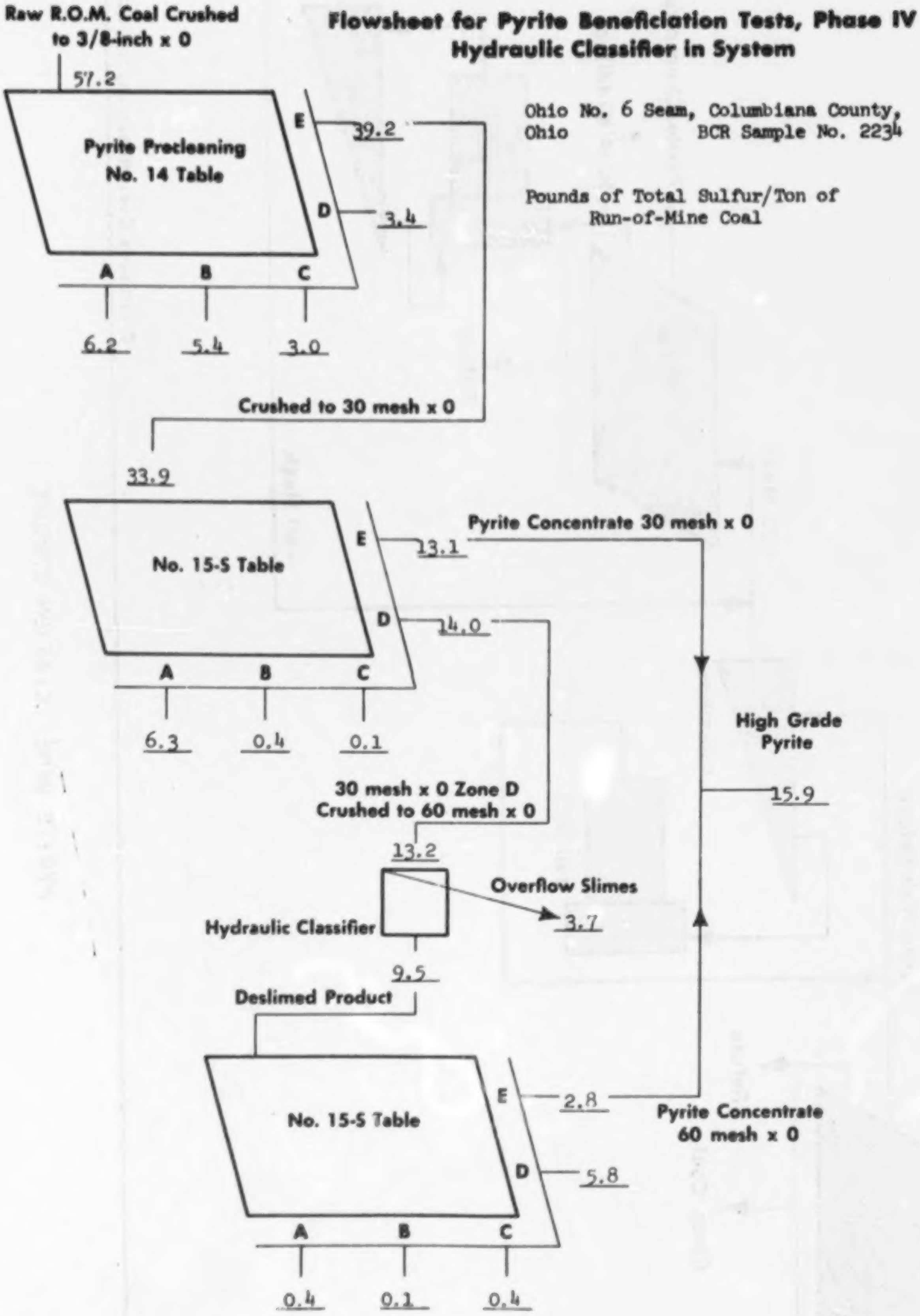

Feed
Sampling Zones
Zone A
Ultra Clean Coal
Chemical Analysis*
Zone B
Clean Coal
Chemical Analysis*
Zone C
High Ash Coal
Float-Sink at 1.60 sp. gr.
Float,
Sink
Chemical Analysis*
Adjustable Division
Zone D
Refuse
Float-Sink at
1.60 and 2.95 sp. gr.
1.60 Float, 1.60 x 2.95, 2.95 Sink
Chemical Analysis*
Zone E
Pyrite
Float-Sink at 2.95 sp. gr.
Float,
Sink
Chemical Analysis*

*Note: Chemical Analysis Consists of Moisture, Ash, and Total Sulfur Plus Ultimate Carbon on 2.95 Sink Fractions

Bituminous Coal Research, Inc. 6070G22

Sampling Flow Sheet for Concentrating Table Tests

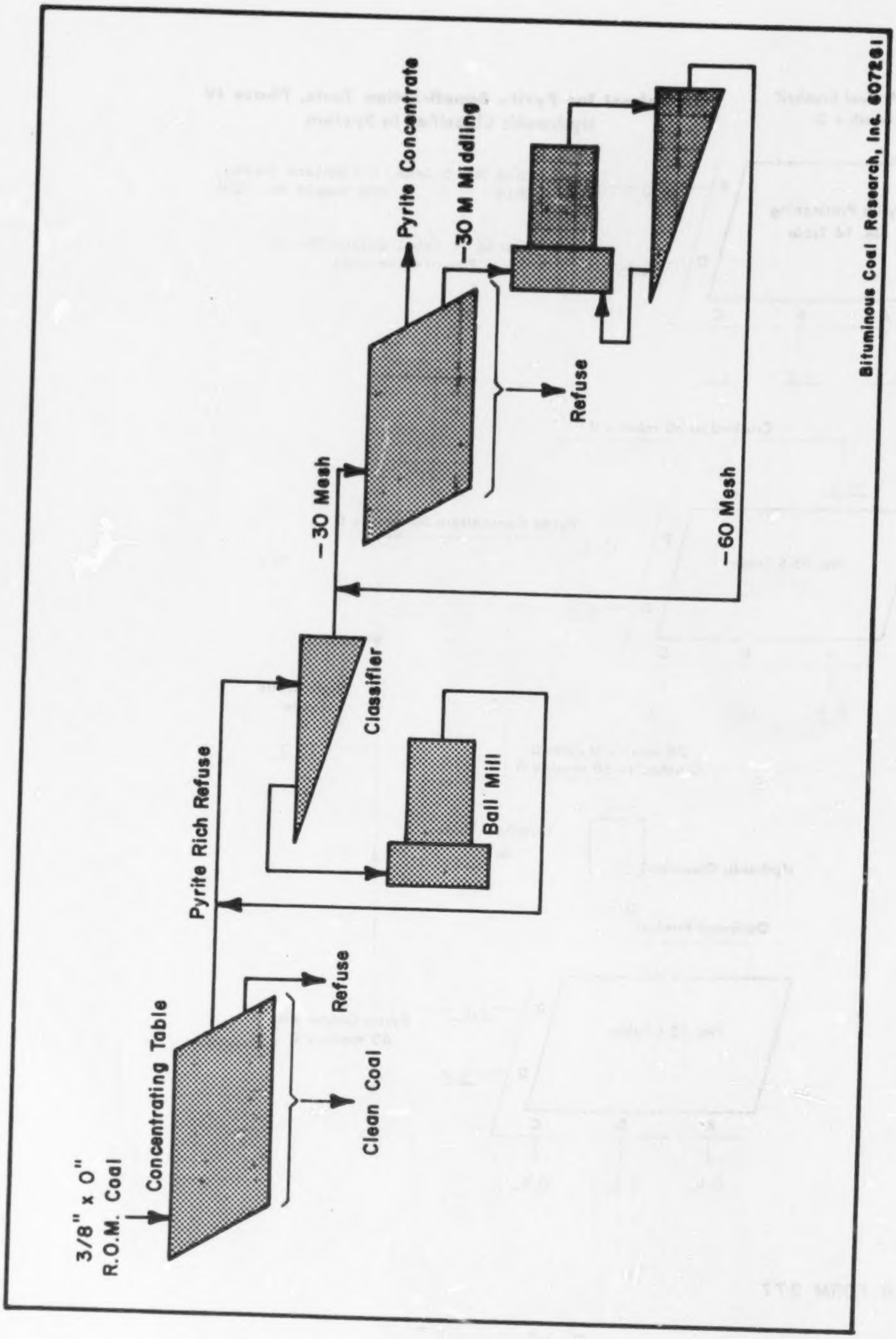
Flowsheet for Pyrite Beneficiation Tests, Phase IV
Hydraulic Classifier in System


Raw R.O.M. Coal Crushed to 3/8-inch x 0
57.2
Pyrite Precleaning No. 14 Table
E
39.2
D
3.4
A
B
C
6.2
5.4
3.0
Crushed to 30 mesh x 0
33.9
No. 15-S Table
E
13.1
D
14.0
A
B
C
6.3
0.4
0.1
Pyrite Concentrate 30 mesh x 0
High Grade Pyrite
15.9
30 mesh x 0 Zone D Crushed to 60 mesh x 0
13.2
Hydraulic Classifier
Overflow Slimes
3.7
9.5
Deslimed Product
No. 15-S Table
E
2.8
D
5.8
A
B
C
0.4
0.1
0.4
Pyrite Concentrate 60 mesh x 0

Ohio No. 6 Seam, Columbiana County, Ohio BCR Sample No. 2234

Pounds of Total Sulfur/Ton of Run-of-Mine Coal

FIGURE II-5


3/8" x 0"
R.O.M. Coal
Concentrating Table
Clean Coal
Refuse
Pyrite Rich Refuse
Classifier
Ball Mill
-30 Mesh
Refuse
Pyrite Concentrate
-30 M Middling
-60 Mesh
Bituminous Coal Research, Inc. 607261

PYRITE BENEFICIATION CIRCUIT

In Tables II-1 through II-5* are entered results of treatments exerted on Sample 2234, Ohio No. 6 Seam. Note that the sulfur concentration in a significant fraction exceeded 48% (91% FeS_2) in the run of 7-28-69, (rerun of Zone E from the first tabling) while the precursor run of 7-1-69 captured 9.5% of the ROM in Zone E at more than 20% sulfur (38.8% FeS_2). BCR Lot No. 1820, from the same mine from which this sample was taken, contained 2.2% total sulfur (Ref. JJ - p. A-120) of which pyritic sulfur accounted for 1.8%. On crushing to minus 30 mesh the pyrite liberation was excellent since 86.4% of the ROM floated at 1.60 specific gravity and contained only 0.5% total sulfur. The sink amounted to 13.6% and contained 15.1% sulfur corresponding to 28.3% FeS_2.

No doubt there are only a limited number of coals from which pyritic sulfur can be so easily and completely separated as from this Ohio No. 6 sample, but the deep cleaning concept and a typical method of accomplishment are sufficiently illustrated by this example.

---

* Ref (JJ)

TABLE II-1

Evaluation of Coal Cleaning Processes and Techniques for Removing Pyritic Sulfur from Fine Coal

Concentrating Table Tests

Coal Identification Ohio No. 6 Seam, Columbiana County, Ohio

Raw Run-of-Mine Coal Crushed to 3/8 Inch x 0, Pyrite Precleaning

BCR Sample No. 2234

Table Products	Product Weight Percent	Float and Sink Weight Percent	Chemical Analysis, Dry Basis, Weight Percent		
			Ash	Total Sulfur	Ultimate Carbon
Zone A	40.2		6.50	0.76	
Zone B	30.3		6.90	0.88	
Zone C					
Float at 1.60	13.6	98.0	7.68	0.98	
Sink at 1.60	0.3	2.0	45.4	4.42	
Composite	13.9	100.0	8.43	1.05	
Zone D					
Float at 1.60	4.34	71.2	15.8	1.84	33.5
1.60 x 2.95	1.72	28.2	44.4	4.60	41.6
Sink at 2.95	0.04	0.6	63.4	26.6	13.9
Composite	6.1	100.0	24.22	2.77	35.67
Zone E					
Float at 2.95	5.9	62.1	65.9	7.42	
Sink at 2.95	3.6	37.9	61.4	42.5	
Composite	9.5	100.0	64.19	20.72	
Composite of Zones A, B, C	84.4		6.96	0.85	
Composite of 1.60 Float Fractions	88.4		7.27	0.89	
Composite of 2.95 Sink Fractions	3.6		61.42	42.33	
Composite of Table Products	100.0		13.45	2.86	
Analysis of Feed to Table			14.8	3.19	

BCR Form No. 139R
10-69
6070

Run Date: 7-1-69

TABLE II-2

Evaluation of Coal Cleaning Processes and Techniques for Removing Pyritic Sulfur from Fine Coal

Concentrating Table No. 15-S Tests - Pyrite Beneficiation

Coal Identification Ohio No. 6 Seam, Columbiana County, Ohio
Zone D, 30 Mesh x 0 Run (7-28-69) Crushed to 60 Mesh x 0

BCR Sample No. 2234

Table Products	Product Weight Percent	Chemical Analysis, Dry Basis, Weight Percent: Ash	Sulfur	Ultimate Carbon
Zone A	36.3	72.4	17.5	
Zone B	15.0	72.8	16.8	
Zone C	7.5	73.2	16.8	
Zone D	34.3	71.2	30.5*	4.4
Zone E	6.9	65.4	48.5	1.9
Composite of Table Products	100.0	71.63	23.94	
Analysis of Feed to Table		72.40	25.40	

* Cut between Zones D and E should have favored zone D more than it did.

BCR Form 172

Run Date: 8-21-69

TABLE II-3

Evaluation of Coal Cleaning Processes and Techniques for Removing Pyritic Sulfur from Fine Coal

Concentrating Table No. 15-S Tests - Pyrite Beneficiation

Coal Identification Ohio No. 6 Seam, Columbiana County, Ohio

Zone E, 3/8 Inch x 0 Run (7-1-69) Crushed to 30 Mesh x 0

BCR Sample No. 2234

Table Products	Product Weight Percent	Chemical Analysis, Dry Basis, Weight Percent: Ash	Total Sulfur	Ultimate Carbon
Zone A	52.9	52.2	6.28	
Zone B	3.2	71.6	6.72	
Zone C	0.5	69.6	14.10	
Zone D	29.2	72.4	25.40	5.4
Zone E	14.2	64.4	48.60	3.9
Composite of Table Products	100.0	60.54	17.93	
Analysis of Feed to Table		64.19	20.72	

BCR From 172

Run Date: 7-28-69

Evaluation of Coal Cleaning Processes and Techniques for Removing Pyritic Sulfur from Fine Coal

Pyrite Beneficiation - Effects of Two-stage Cleaning

Coal Identification OHIO NO. 6, COLUMBIANA COUNTY, OHIO

BCR Sample No. 2234

Pyrite Precleaning

Concentrating Table No. 14 Test Run Date 7-1-69
Feed to Concentrating Table: Raw Run-of-Mine Coal Crushed to 3/8 Inch x 0

	Product, Weight %	Chemical Analysis, Ash	Dry Basis, Weight % Total Sulfur
Analysis of Feed to Table	100.0	14.80	3.19
Zone E (Pyrite Zone)	9.5	64.19	20.72

Pyrite Cleaning

Concentrating Table No. 15-S Test Run Date 7-28-69
Feed to Concentrating Table: Zone E, 3/8 Inch x 0 Run (7-1-69) Crushed to 30 Mesh x 0

	Product, Weight %	Chemical Analysis, Ash	Dry Basis, Weight % Total Sulfur
Analysis of Feed to Table	100.0	64.19	20.72
Zone E (Pyrite Zone)	14.2	64.40	48.60

Concentrating Table No. 15-S Test Run Date 8-21-69
Feed to Concentrating Table: Zone D, 30 Mesh x 0 Run (7-28-69) Crushed to 60 Mesh x 0, Classifier Product

	Product, Weight %	Chemical Analysis, Ash	Dry Basis, Weight % Total Sulfur
Analysis of Feed to Table	100.0	71.80	26.40
Zone E (Pyrite Zone)	16.9	66.20	46.80
Two Stage Pyrite Product	1.6	64.70	48.30

BCR FORM 173

TABLE II-5

Evaluation of Coal Cleaning Processes and Techniques for Removing Pyritic Sulfur from Fine Coal

Flowsheet Data for Pyrite Beneficiation Tests

Coal Identification: Ohio No. 6 Seam, Columbiana County, Ohio

BCR Sample No. 2234

Fraction	Weight Percent		Chemical Analysis, Dry Basis Weight Percent	
			Ash	Total Sulfur
Model No. 14 Concentrating Table				
Zone A	40.2		6.50	0.76
Zone B	30.3		6.90	0.88
Zone C	13.9		8.43	1.05
Zone D	6.1		24.22	2.77
Zone E		9.5	64.19	20.72
Model No. 15-s Concentrating Table				
Zone A	5.0		52.2	6.28
Zone B	0.3		71.6	6.72
Zone C	0.1		69.6	14.10
Zone D		2.8	72.4	25.40
Zone E	1.3		64.4	48.60
Model No. 15-s Concentrating Table				
Zone A	1.0		72.4	17.5
Zone B	0.4		72.8	16.8
Zone C	0.2		73.2	16.8
Zone D	1.0		71.2	30.5
Zone E	0.2		65.4	48.5
Composite of Fractions	100.0		13.09	2.54
Analysis of Feed Coal			14.8	3.19

7. PROCESS CONTROL OF REJECT STREAMS

It is clear that cleaning circuits such as illustrated in Figure II-4 and II-5 can effect good separation of coal and liberated pyrite. This capability extends to circuits of other devices and combinations of devices designed to bring about gravimetric separations. Both coal and pyrite can be separated from liberated ash substances although less efficiently as specific gravity differentials become smaller. From the technical point of view, the key word is liberation — which can be promoted, if at all, only by size reduction (especially by crushing and grinding). Refer again to Figure II-1 in which summations of washing tests on 70 samples show a clear improvement tendency in pyritic sulfur reduction as a function of smaller sizing of the ROM.

7.1 Thus, if size reduction will adequately liberate the several fractions of a ROM, it is feasible:

a) To separate the fractions to recover a specification coal low in pyritic sulfur, and

b) To recover pyritic fractions of low coal content, and

c) By recombining b) with some of a) produce a high sulfur fuel having a controlled ratio of pyritic sulfur to coal, that is, a "tailor-made" or custom specification fuel.

8. Thus, in general, any ROM coal containing pyritic sulfur that can be processed to yield coal of lower sulfur content will yield at the same time a fuel of higher sulfur content in which the sulfur/coal ratio can be controlled to any specification within wide limits. This is the central

basis of fact in the present study. The remaining question (limiting the point of view of the requirements of this study) is the one of cost. Put another way, this question is:

> What is the value of the fuel having the higher sulfur to coal ratio?

But, no general answer to this question, such as would be provided by a broad market, can be expected. The question is meaningful only in a specific context in which the fuel is processed and the energy and sulfur products sold. Estimates of the value of such fuels, reflecting specific processing and marketing assumptions, are included in Part VIII of this study.

III

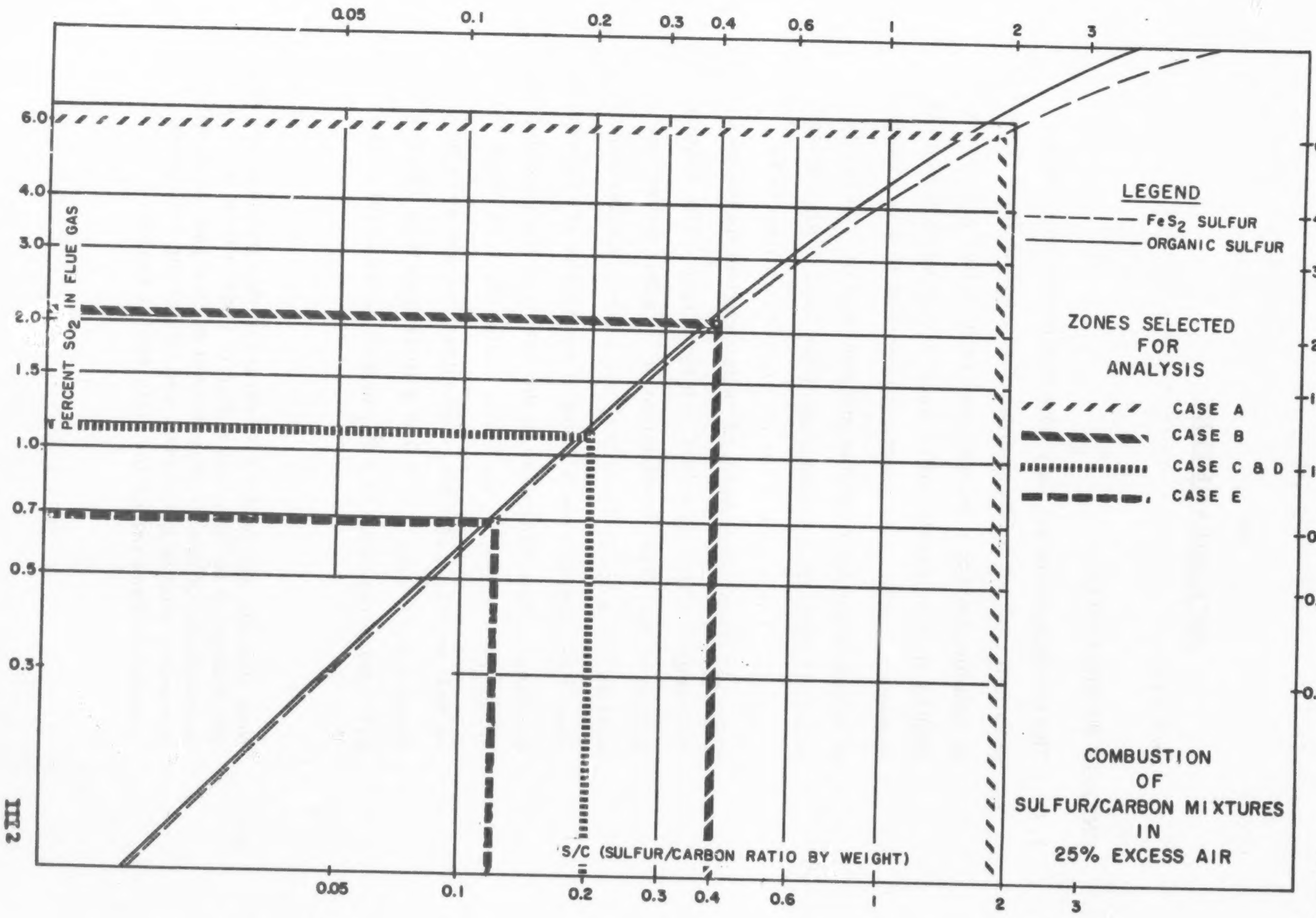
BASE PARAMETERS

1. ENERGY EXTRACTION

1.1 The combustion in air (say 25% excess air which is a parameter of the boiler designs discussed in Part IV and in Volume Two) of a known mixture of sulfur and carbon results in predictable sulfur dioxide concentration in the combustion gas. For practical purposes this is also true of known mixtures of sulfur and coal and of iron pyrite and coal when the composition of the coal is also known. Figure III-1 is a plot of the SO_2 concentrations to be expected from combustion of sulfur/carbon mixtures in the range S/C = 0.02 to S/C = 6 by weight.* The concentration of SO_2 from combustion of "organic sulfur" in coal differs slightly from SO_2 concentrations developed from FeS_2 (pyrite) combustion if conditions are otherwise the same. The difference is the result of the consumption of oxygen to form, in the latter instance, iron oxide ash as well as SO_2. The SO_2 concentrations can span the range from 0% for S/C = 0 to a little less than 13% for $S/C = \infty$ (combustion of FeS_2 in 25% excess air). The

---

* Note that the use of the S/C ratio for this purpose avoids the ambiguity of "percent sulfur in coal" which is determinant of total sulfur but not of flue gas composition. However, coal is not carbon and use of the S/C ratio results in some minor inconsistencies herein.



COMBUSTION
OF
SULFUR/CARBON MIXTURES
IN
25% EXCESS AIR

FIGURE III-1

curves of Figure III-1 show only the middle portion of this range since interest here is limited to combustion of sulfur and coal in mixtures, say, between the limits S/C = 0.08 and S/C = 2. These limits will yield SO_2 concentrations higher than result from burning "high sulfur" coal and lower than result from roasting iron pyrite.

The rationale of this limitation is simply that since the HSC Fuels can be "tailor-made" by the cleaning plant (as argued in Part II) the specification should be cut to serve the downstream processes, cost considered. Since these processes will be designed and operated for economical production of valuable heat and sulfur products, the HSC Fuel should be of reasonably high heating value consistent with a substantial sulfur content.

1.2 In respect of heating value it is noted that combustion of iron pyrite yields about 3,000 BTU per pound while a low ash bituminous coal may be 14,000 BTU per pound or more.

Generally speaking, the available technology in the SO_2 recovery arts is benefitted by high rather than low flue gas concentrations of SO_2. This is true mainly because gas purification equipment is more costly if larger rather than smaller volumes of gas need to be handled.

1.3 These considerations, and others later discussed, are the basis for confining this exploration within boundaries as follows:

	HSC FUEL	
	Low Sulfur	High Sulfur
S/C Ratio	0.08	2
Percent SO_2 in Flue	0.4 - 0.5	6 - 7
On Non-Pyritic-Ash-Free Basis:		
Percent S in fuel, ca.	7	42
Equivalent percent FeS_2 in fuel, ca.	13	79
Percent coal combustibles* in fuel, ca.	87	21
Higher heating value of fuel Btu/Pound, ca.	12,600	5,300

* Sulfur-free

It is believed that a coal cleaning plant can be designed to produce low sulfur coal plus one or more high sulfur fuels within the range of compositions defined above. But this is a broad range and any processing system of the types considered herein will operate well only on a narrow range of fuel composition. That is to say fuel held to a reasonably narrow specification that must be one of the principal parameters of the process design. The process systems will not, however, be designed around the fuel specifications as is the normal procedure for fossil fuel furnaces. The logic here is to design the process to fit the market and to tailor-make the fuel to fit the process.

Available technology does not appear to include a single type of process system that will do well enough with the diversity of possible market situations that will dominate the recovered sulfur value, and offer at the same time the design flexibility required for adaptation to output of 500,000 lbs/hr of steam at one location or 500 MW at another. (See Statement of Contract Work, p I-2).

1.4 Three process systems of presently available technology appear to be no more than necessary to exploit all zones of the opportunity that may be offered by the high sulfur combustor thesis. These three have been developed into five case studies* as follows:

1.4.1 CASE A system is designed for firing an HSC Fuel having an S/C ratio of about 2 to produce 500,000 lbs/hr of industrial quality steam and a flue gas of 6% SO_2 concentration which is directly converted to some 1,800 tons/day of commercial grade sulfuric acid. These outputs would be secondary coproducts of clean coal produced at a rate of 7 million tons per year by washing a run of mine coal containing 3% pyritic sulfur. The actual clean coal output would, of course, depend on the actual ROM composition and washability.

A system of this type is conceived to be of possible value as an addition to an existing coal-burning utility complex (See 1.6.2), or existing chemical or steel-making industry (See 1.6.3).

1.4.2 CASE B system will fire an HSC Fuel of about 0.4 S/C ratio (2% SO_2 in flue gas). It is scaled to 500,000 lbs/hour industrial steam as in CASE A, but much smaller outputs of clean coal and sulfuric acid would be co-produced.

---

* See Part I Abstract, or Table IV-7, for tabulated summary of the five case studies.

1.4.3 CASE E employs SO_2 recovery technology (Monsanto Cat-Ox) similar to CASE B but this system is scaled to the 500 MW output level (more than 10 times the energy capacity of CASE B) so that the combustion equipment is quite different. The SO_2 recovery trains operate on relatively weak gas of 0.7% SO_2 which is generated by an S/C ratio of 0.12.

1.4.4 CASE D is also scaled to 500 MW energy output but the SO_2 recovery technology is quite versatile compared to processes mentioned above. The acid intermediates: liquid SO_2, magnesium sulfite or sulfur, can be produced for storage or shipment, or sulfur value can be converted directly to sulfuric acid, any of these by available technology. The main design theme of CASE D in this study is for output of sulfur at 770 net tons per day from a 1% SO_2 flue gas that requires an S/C ratio of 0.2.

1.4.5 CASE C is the reduced-scale prototype or demonstration unit of CASE D, designed to operate on the same HSC Fuel composition but at 50 MW energy capacity and about 100 tons per day of sulfur.

Thus the five cases include two of industrial interest, two of electric utility interest, and a prototype unit in the domain of the latter.

a

HIGH SULFUR COMBUSTOR

COAL, ENERGY & SULFUR P

MILLION TONS/YEAR

CASE	RUN OF MINE	CLEAN COAL	HSC FUEL
	(DRY BASIS)		
D	9.7	7.1	2.3
A	8.1	7.1	0.64
E	5.8	3.5	2.1
B	2.8	2.3	0.41
C	1.3	0.92	0.30

NOTES

(1) ROM: 1% ORGANIC S, 3% PYRITIC S
13% NON-PYRITIC ASH.

(2) CLEAN COAL: 1% ORGANIC S
5% NON-PYRITIC ASH

(3) CLEAN COAL + HSC FUEL=95% OF ROM WEIGHT (DRY BASIS)

(4) TOTAL CLEANING LOSSES = 5%

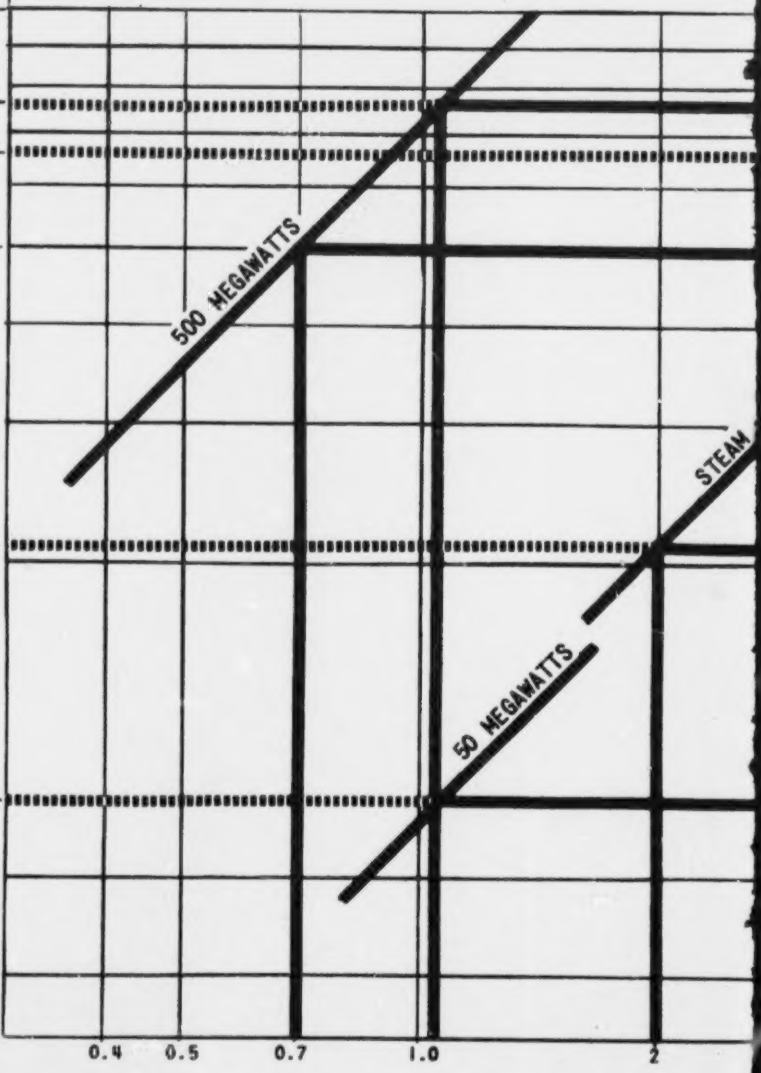
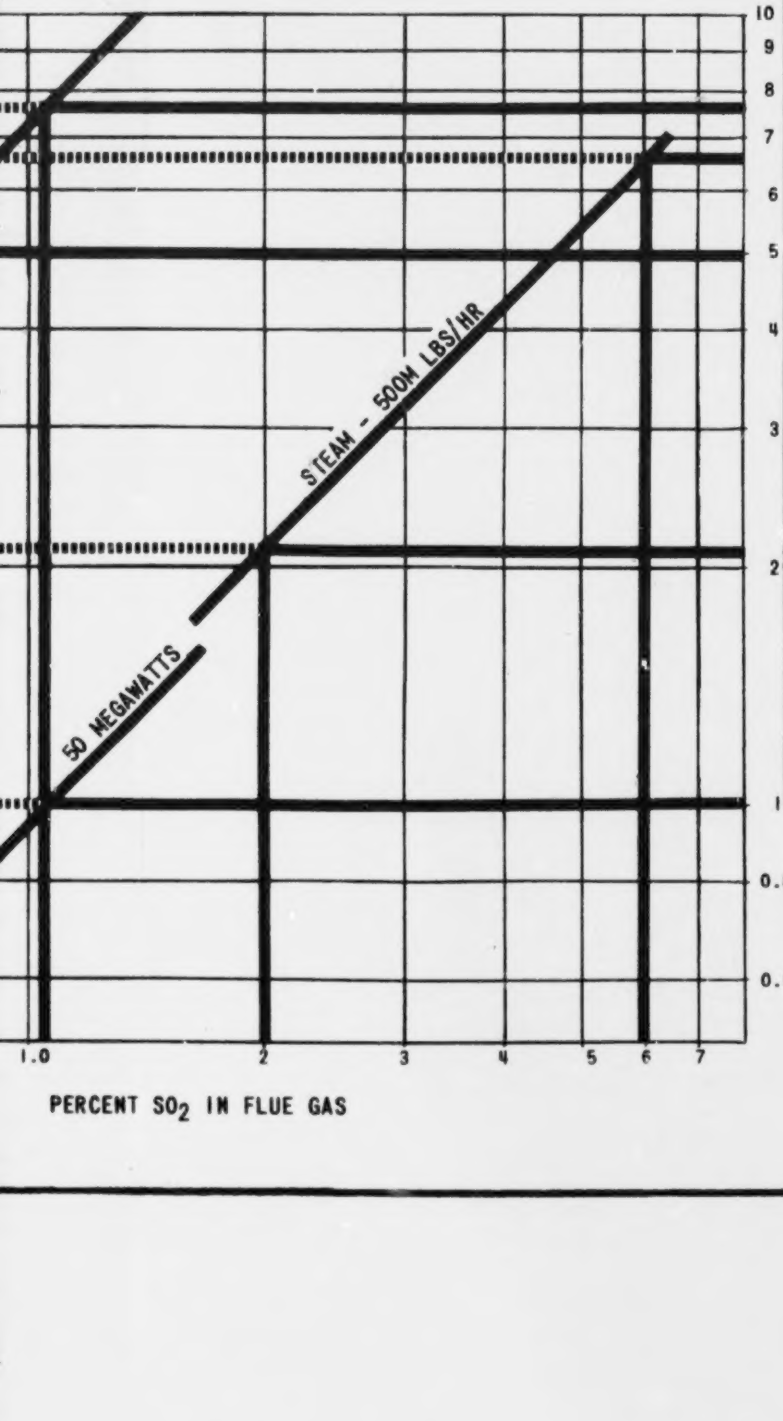


FIGURE III-2

HIGH SULFUR COMBUSTOR STUDY

OAL, ENERGY & SULFUR PRODUCTS



NET TONS/DAY				
SULFUR	SULFURIC ACID EQUIVALENT	MAGNESIUM SULFITE EQUIVALENT	LIQUID SULFUR DIOXIDE EQUIVALENT	CASE
760⊛	2300	2500⊛	1520	D
640	1830⊛	1980	1220	A
510	1420⊛	1580	970	E
230	620⊛	710	440	B
100⊛	291	310⊛	190	C

⊛ DESIGN BASIS IN THIS STUDY

The HSC Fuel and consequent flue gas compositions are indicated in Figure III-1, and with these as base parameters the diagram of Figure III-2 shows the main related quantities. The various fuel tonnages entered in the left hand columns, and the sulfur value quantities in the right hand columns, are related by the assumption that the ROM contains 3% pyritic sulfur and 1% organic sulfur, that the pyritic sulfur would wash out completely, and that the non-pyritic ash distribution would be as indicated in the NOTES on the diagram.

It is clear, of course, that any change in fuel compositions will affect all quantities. The whole diagram represents approximate arithmetic relationships that are adjusted for expected operating efficiencies in combustion and SO_2 recovery. The yields of pyritic sulfur and carbonaceous material from the washing operation may be given effect in the diagram at levels that are too high for any actual practice. These yields will depend on washabilities that vary widely, of course, from one coal to another. Yield of HSC Fuel and total heat value yield will depend particularly on efficiency of non-pyritic ash separation. Since the combustion products of the HSC Fuels are exceptionally high in pyritic ash, the possibility of loading up also with non-pyritic ash is not unlimited. The subject is discussed in Parts IV and XI.

1.5 The diagram of Figure III-2 does illustrate the heating value effect of the varying proportions in the HSC Fuels of the low heating value component (FeS_2) and the high value component (coal). CASES D and A for example, based on entirely different technology, are intended to produce respectively 500 MW and 500,000 lbs/hour of steam (about 50 MW) - a ten-fold difference in energy output, while the yield of sulfur value is similar as to quantity, and the yield of "clean coal" a little over 7 million tons per year from each. The choice here would evidently depend on the market for energy. If a very large electric capacity is wanted, say 1,000 MW, the CASE D or E system would be designed to operate on a lean flue gas of 0.5% SO_2 or less in order to avoid overproduction of sulfur value at one location, and in order to realize a high output of clean coal at the same time. If a low energy capacity will satisfy local demand, a combustor and sulfur recovery system such as illustrated by CASE A may suffice.

It is not overlooked that the 7 million ton clean coal magnitude may be too large to expect from one cleaning plant since even the largest single bituminous mines are well under this level of annual output. Several sources would probably have to contribute to such a volume of coal. The diagram of Figure III-2 indicates for example that clean coal at about 3 million tons per year might be associated with the HSC Fuel for a 500 MW capacity, a flue gas of 0.4% SO_2 and 250 tons per day of sulfur value as sulfur.

2. SULFUR VALUE

2.1 Sulfuric Acid

In this study it is assumed that the only important market for "sulfur value" extracted from coal reject material is the one dominated by the sulfuric acid consumer. The sulfuric acid end-product, however, may be manufactured either directly from the sulfur dioxide in the flue gas of the high sulfur combustor, or manufactured from any of three (possibly more) intermediate materials, producible from the flue gas, each of which can be stored or shipped, each on terms peculiar to it.

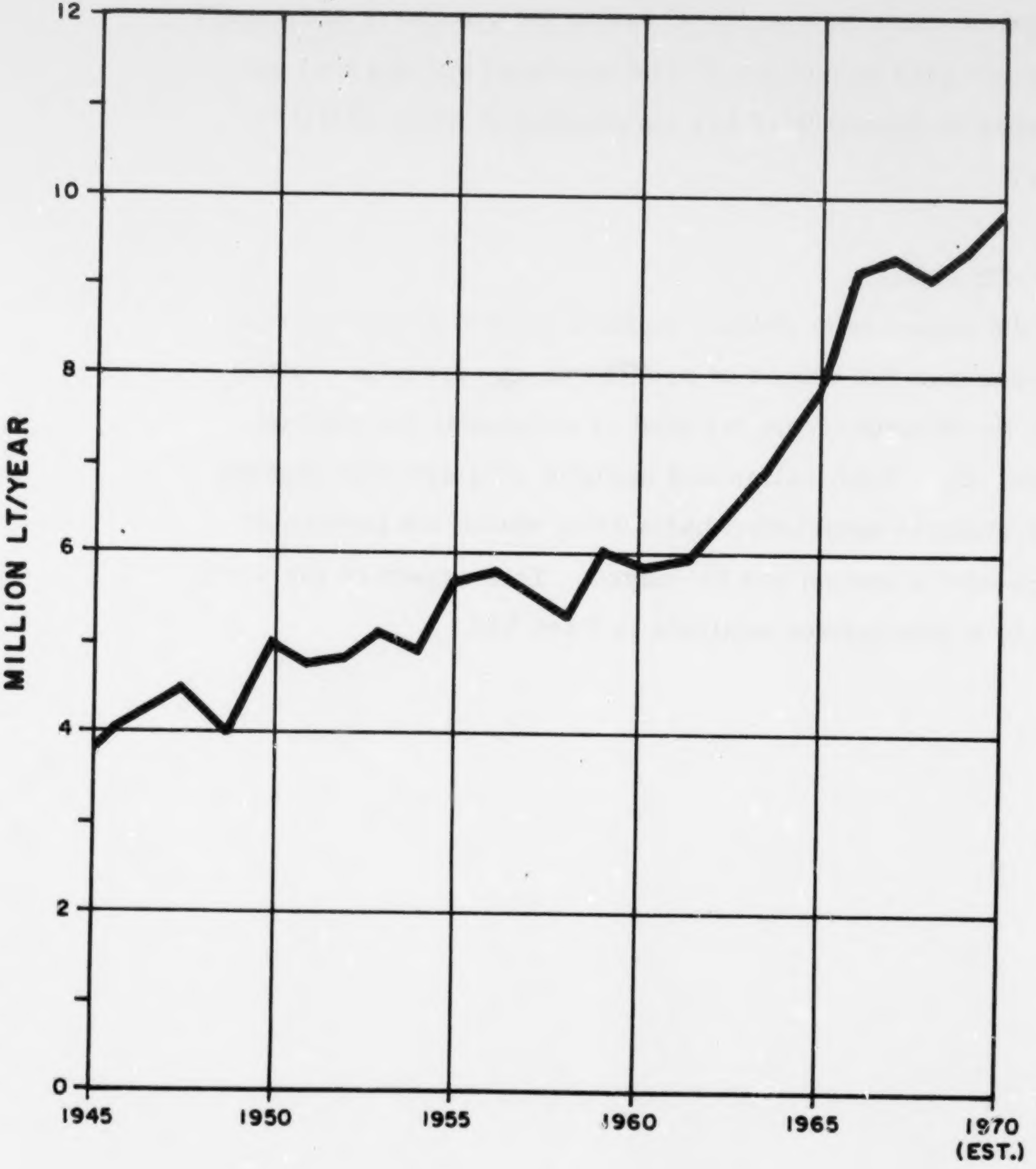
2.2 Sulfur

By far the most widely employed intermediate (ordinarily called "raw material") is commercial grade sulfur of which annual world production is now more than 35 million long tons of which 9 to 10 million long tons is produced by mining or off-product recovery in the United States. (Figure III-3). Since one ton of sulfur will make about three tons of acid, shipment of sulfur where possible is much preferred to shipment of acid. Sulfur moreover can be stored uncovered on the ground to meet seasonal fluctuations in demand, as acid cannot. For these and other reasons sulfur enjoys great marketing flexibility compared to the end-product, sulfuric acid; the SO_2 recovery process of CASE D and its prototype CASE C is thus designed primarily for production of commercial grade sulfur.

2.3 Magnesium Sulfite and Liquid Sulfur Dioxide

However, the CASE D system yields an intermediate, dry magnesium sulfite, that can be shipped to another processing point, but whether shipped or not this material can be converted to sulfuric acid as well as sulfur, or to liquid sulfur dioxide if wanted, since the latter substance may also be converted to

FIGURE III-3



CONSUMPTION OF SULFUR
IN THE
UNITED STATES

REF. (EE)

sulfuric acid. CASE A system will produce sulfuric acid only, and only from a high S/C ratio fuel; CASE B and CASE E systems function on lower S/C ratios but similarly are capable of sulfuric acid output only. The technical options that are suggested in Figure III-2 are considered in more detail in Part VI.

2.4 Trade-Off Basis

Since the object is to deliver sulfuric acid to a customer at minimum cost the choice of process in any specific context cannot be detached from the cost of shipment, the cost of storage, etc. Both sulfur and sulfuric acid are high-lighted in this study to establish a basis from which the particular and specific situation can be judged. This aspect of the study is put to a generalized analysis in Part VIII.

IV
FIRING OF HSC FUELS

1. FUEL COMPOSITIONS

The development of the HSC Fuel Specifications is discussed in Volume Two Part XI and summarized briefly here. First, it is taken for granted that the specifications should be compatible to whatever degree possible with optimum total yield of fuel value from the cleaning plant. Second, however, the S/C ratio of the HSC Fuel must be held to narrow limits to stabilize the heating value of the fuel and to avoid penalizing the sulfur recovery operation; and third, the non-pyritic ash content must be controlled, in part to maintain specific heating value but more importantly to avoid an unpredictable ash fusion temperature. These are major considerations.

1.1 Whole Refuse Fraction

Maximum cleaning plant yield would result from extraction of heat and sulfur value from the entire refuse fraction without waste. While this possibility is unlikely to occur frequently, it is not unknown. A review of some washability tests in current programs has identified five samples taken from seams containing substantial coal reserves that were crushed to 1.5 inch by 100 mesh and separated at ± 1.35 specific gravity. The two fractions of each sample were analyzed with results as shown in Table IV-1.

TABLE IV-1

WHOLE REFUSE FRACTION FROM FIVE ROM COALS

Coal #	Clean Coals		REFUSE COMPOSITIONS					
	Weight %	Total S%	Weight %	Pyritic Sulfur %	Total S%	Non-Pyritic Ash %	Coal Combustibles %	Sulfur/Coal Ratio
1	90	2.0	10	10.2	14.0	21.9	59.0	0.24
2	70	2.0	30	11.0	14.0	28.4	51.0	0.27
3	75	2.0	25	10.0	12.7	23.9	57.3	0.22
4	70	<1.0	30	10.7	13.5	28.8	51.1	0.26
5	75	2.5	25	9.3	12.7	14.3	68.3	0.19

LOCATION

Coal #	State	County	Seam
1	Ohio,	Muskingum	No. 6
2	Ohio,	Tuscarawas	No. 6
3	Pennsylvania,	Indiana	Lower Kittanning
4	Pennsylvania,	Indiana	Lower Kittanning
5	Ohio,	Muskingum	No. 5

NOTE: The spread between Pyritic Sulfur and Total Sulfur in the refuse compositions suggests that the concentration of Organic Sulfur in the "Coal Combustible" component of the Refuse is substantially greater than it is in the Clean Coal. If so, here is evidence of some gravimetric separation of organic sulfur along with the pyritic sulfur.

A very minor adjustment of these refuse or reject compositions would produce the HSC Fuel required for the combustor system of CASE C or D.

1.2 Blending of Fractions

More often however, it is expected that the reject material will need to be cut into two or more fractions of which portions will be blended, possibly with the addition of some clean coal, to produce specification HSC Fuels.

1.3 Heat Value Yield and Non-Pyritic Ash

Whatever the blending or adjusting procedure, the content of non-pyritic ash will be a critical factor in the total heat value yield of the cleaning plant and in the success of the HSC Fuel itself in serving its intended purpose. The reasons are briefly:

a) High yield in cleaning operations depends on the extent of utilization of middlings, that is coal from which ash failed to separate. While high total yield of heat value and optimum economic results are by no means the same thing, low yield will certainly result in bad economics. The usual incentive will be to load up the HSC Fuels with as much ash as possible in order to save the associated coal.

b) Since however the HSC Fuels are high ash anyway, as a result of the inescapable iron oxide, 1.25 pounds of it per pound of pyritic sulfur in the fuel, the loading of non-pyritic ash (that is the component of the total ash of the fuels that does not derive from iron pyrite) must be limited by the capability of the furnace equipment designed to burn it.

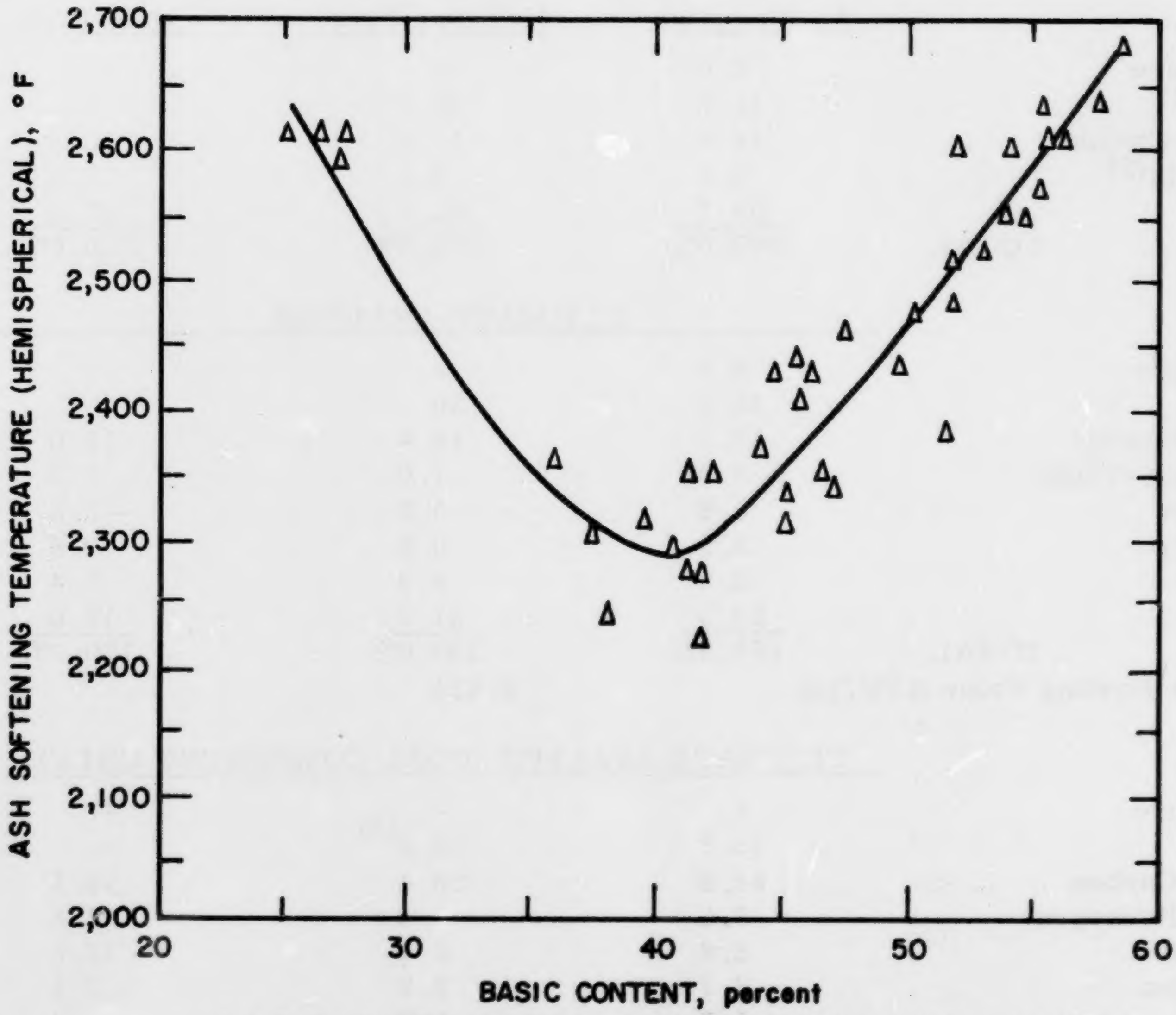
The composition as well as the quantity of non-pyritic ash, of course, matters of some complexity, are discussed in Part XI in which theoretical and experimental criteria are analyzed in order to establish the bases of the HSC Fuel Specficiations. Fuel volatility, reactivity, ignition temperature, grindability, corrosion and erosion are also among the critical properties considered in the design of combustion equipment for the HSC systems.

1.4 Ash Fusion Temperature

Of special interest are ash fusion characteristics which are, in the HSC Fuels, expected to be much influenced by the unusually high concentration of iron, a basic constituent. The effect of basicity is plotted in Figure IV-1. The HSC Fuel compositions are intended to yield total ash having softening temperatures rather high on the right hand side of this curve, i.e., 50% "basic content" or greater. Among the important specified limitations on non-pyritic ash is that it should not exceed 20 pounds per 100 pounds of dry HSC Fuel and that alkali metal oxides, calculated as Na_2O, should not exceed four tenths pounds of the twenty. See Part XI.

1.5 Four HSC Fuels are specified - one each for CASE A, B and E, and one for CASES C and D. The specifications are in Table IV-2 through IV-5 in terms of the Proximate and Ultimate Analyses of the HSC Fuel and the Ultimate Analysis of the coal component.

FIGURE IV-1



Ash Softening Temperature Depending on Basic Content of Coal Ash.

REF. (See Part XI)

TABLE IV-2

HSC FUEL COMPOSITION

CASE A

	PROXIMATE ANALYSIS		
	As Received	Moisture Free	Moisture Free Ash Free
Moisture	6.0	--	--
Ash (1)	18.8	20.0	--
Fixed Carbon	10.3	11.0	13.8
Volatile(2)	6.8	7.3	9.2
FeS_2	58.1	61.7	77.0
TOTAL	100.0%	100.0%	100.0%
	ULTIMATE ANALYSIS		
Moisture	6.0	--	--
Ash	18.8	20	--
Total Carbon	13.5	14.4	18.0
Total Hydrogen	0.9	1.0	1.2
Oxygen	2.0	2.2	2.8
Nitrogen	0.5	0.5	0.6
Sulfur	0.3	0.3	0.4
FeS_2	58.0	61.0	77.0
TOTAL	100.0%	100.0%	100.0%
Higher Heating Value BTU/LB.		4,424	
	ULTIMATE ANALYSIS (COAL COMPONENT ONLY)		
Moisture	6	--	--
Ash	23.7	25.2(3)	--
Total Carbon	54.8	58.4	78.5
Total Hydrogen	3.6	3.8	4.8
Oxygen	8.8	9.4	12.5
Nitrogen	2.1	2.2	2.8
Sulfur	1.0	1.0	1.4
TOTAL	100.0%	100.0%	100.0%

---

(1) "Ash" in this tabulation means non-pyritic ash. Ash specification is discussed in Part XI.

(2) Volatile matter as derived from the coal component. One of the S atoms in the FeS_2 is also "volatile" (labile) above about 750° F. Total Volatility may reach 30% of total combustibles in this fuel.

(3) ROM coal may have only 5-10% non-pyritic ash, which may tend to concentrate in the HSC Fuel.

TABLE IV-3

HSC FUEL COMPOSITION

CASE B

	PROXIMATE ANALYSIS		
	As Received	Moisture Free	Moisture Free Ash Free
Moisture	6.0	--	--
Ash(1)	18.8	20.0	--
Fixed Carbon	26.1	27.8	34.8
Volatile(2)	17.3	18.4	23.0
FeS_2	31.8	33.8	42.2
TOTAL	100.0%	100.0%	100.0%

	ULTIMATE ANALYSIS		
Moisture	6.0	--	--
Ash	18.8	20.0	--
Total Carbon	34.1	36.3	45.4
Total Hydrogen	2.1	2.2	2.8
Oxygen	5.5	5.8	7.3
Nitrogen	1.2	1.3	1.6
Sulfur	0.5	0.6	0.7
FeS_2	31.8	33.8	42.2
TOTAL	100.0%	100.0%	100.0%
Higher Heating Value BTU/LB.		7,400	

	ULTIMATE ANALYSIS (COAL COMPONENT ONLY)		
Moisture	6	--	--
Ash	23.7	25.2(3)	--
Total Carbon	54.8	58.4	78.5
Total Hydrogen	3.6	3.8	4.8
Oxygen	8.8	9.4	12.5
Nitrogen	2.1	2.2	2.8
Sulfur	1.0	1.0	1.4
TOTAL	100.0%	100.0%	100.0%

(1) "Ash" in this tabulation means non-pyritic ash. Ash specification is discussed in Part XI.

(2) Volatile matter as derived from the coal component. One of the S atoms in the FeS_2 is also "volatile" (labile) above about 750° F. Total volatility may reach 34% of total combustibles in this fuel.

(3) ROM coal may have only 5-10% non-pyritic ash, which may tend to concentrate in the HSC Fuel.

TABLE IV-4

HSC FUEL COMPOSITION

CASE C AND D

		As Received	Moisture Free	Moisture Free Ash Free
		PROXIMATE ANALYSIS		
Moisture		6.0	--	--
Ash$^{(1)}$		18.9	20	--
Fixed Carbon		32.6	36.4	44.2
Volatile$^{(2)}$		21.7	23.5	29.4
FeS$_2$		19.5	21.1	26.4
	TOTAL	100.0%	100.0%	100.0%
		ULTIMATE ANALYSIS		
Moisture		6.0	--	--
Ash		18.9	20	--
Total Carbon		43.0	46	57.5
Total Hydrogen		2.9	3.0	3.7
Oxygen		7.0	7.4	9.2
Nitrogen		1.6	1.7	2.2
Sulfur		.8	.8	1.0
FeS$_2$		19.8	21.1	26.4
	TOTAL	100.0%	100.0%	100.0%
Higher Heating Value BTU/LB.			9,000	
		UTLIMATE ANALYSIS (COAL COMPONENT ONLY)		
Moisture		6.0	--	--
Ash		23.7	25.2$^{(3)}$	--
Total Carbon		54.8	58.4	78.5
Total Hydrogen		3.6	3.8	4.8
Oxygen		8.8	9.4	12.5
Nitrogen		2.1	2.2	2.8
Sulfur		1.0	1.0	1.4
	TOTAL	100.0%	100.0%	100.0%

(1) "Ash" in this tabulation means non-pyritic ash. Ash specification is discussed in Part XI.

(2) Volatile matter as derived from the coal component. One of the S atoms in the FeS$_2$ is also "volatile" (labile) above about 750° F. Total volatility may reach 36% of total combustibles in this fuel.

(3) ROM coal may have only 5-10% non-pyritic ash, which may tend to concentrate in the HSC Fuel.

TABLE IV-5

HSC FUEL COMPOSITION

CASE E

	PROXIMATE ANALYSIS		
	As Received	Moisture Free	Moisture Free Ash Free
Moisture	6.0	--	--
Ash(1)	18.8	20.0	--
Fixed Carbon	37.7	40.1	50.1
Volatile(2)	25.1	26.7	33.4
FeS_2	12.4	13.2	16.5
TOTAL	100.0%	100.0%	100.0%

	ULTIMATE ANALYSIS		
Moisture	6.0	--	--
Ash	18.8	20	--
Total Carbon	49.3	52.4	65.5
Total Hydrogen	3.0	3.2	4.0
Oxygen	7.9	8.4	10.5
Nitrogen	1.8	1.9	2.4
Sulfur	0.8	0.9	1.1
FeS_2	12.4	13.2	16.5
TOTAL	100.0%	100.0%	100.0%
Higher Heating Value BTU/LB.		9,660	

	ULTIMATE ANALYSIS (COAL COMPONENT ONLY)		
Moisture	6	--	--
Ash	23.7	25.2(3)	--
Total Carbon	54.8	58.4	78.5
Total Hydrogen	3.6	3.8	4.8
Oxygen	8.8	9.4	4.8
Nitrogen	2.1	2.2	2.8
Sulfur	1.0	1.0	1.4
TOTAL	100.0%	100.0%	100.0%

(1) "Ash" in this tabulation means non-pyritic ash. Ash specification is is discussed in Part XI.

(2) Volatile matter as derived from the coal component. One of the S atoms in the FeS_2 is also "volatile" (labile) above about 750° F. Total volatility may reach 38% of total combustibles in this fuel.

(3) ROM coal may have only 5-10% non-pyritic ash, which may tend to concentrate in the HSC Fuel.

2. FURNACE TYPES

At this point consideration is given to available combustion equipment with a view to reduction in the number of possible choices on grounds easily established. Well recognized modern furnace types are as follows:

2.1 Roasting Equipment

- a) Kiln
- b) Multiple Hearth
- c) Flash Roasters

2.2 Fluidized Bed Combustors

2.3 Stoker Furnaces

2.4 Grate Furnaces

2.5 Crushed Coal-Fired

- a) Cyclone Furnace

2.6 Pulverized-Fuel Fired

- a) Wet bottom furnace
- b) Dry bottom furnace
    - Horizontally fired
    - Vertically fired

3. FURNACE LOAD REQUIREMENTS

The primary duty of the furnace or combustor will be production of steam of high competitive quality for large volume requirements. This is to say that steam output intended for industrial consumption will be economical for all except specialized uses, and that steam for electricity will efficiently drive large modern turbo-generators.

3.1 Steam Design

The steam design bases are as follows:

Type of Boiler	Temp. °F	Press. PSIG	Quantity M lbs/hr.
Industrial	825	900	500
Utility (50 MW)	1,000	1,450	500
Utility (500 MW)	1,005	2,500	3,500

3.2 Service Factor

In keeping with the primary objective which is production of clean coal and production of more rather than less, the industrial and utility boilers are designed to be base loaded as measured by a service factor of 8,000 hours of full capacity operation per year.

4. FURNACE SELECTION BASIS

The steam production requirement alone appears to eliminate furnace types 2.1, 2.3 and 2.4 on page IV-6. None of these designs can be fired at the necessary high temperatures. Fluidized bed combustion may have a potential capability to produce this quality and volume of steam, but it is not yet established commercially and since certain other factors appear also to be unfavorable as discussed in Part XI, fluidized bed combustion is not considered further in this summary section.

Cyclone furnaces while well developed are notable for high maintenance costs which could be aggravated by the high concentration of iron and sulfur in the HSC Fuel.

Suspension-fired furnaces, however, with either vertical or horizontal burners are not objectionable on any of the foregoing grounds, and enjoy also a long history of commercial success in both dry and wet bottom designs.

Even so, a considerable study of the expected characteristics of the HSC Fuels has been necessary to establish pulverized fuel firing as most suitable and to develop the basis for selection of the vertical or horizontal firing technique as a function of the pyritic content of the HSC Fuels.

In an earlier section of Part IV it is argued that ash fusion temperatures of the HSC Fuels must be stabilized in a predictable range, and, in view of high iron concentrations, this range should be higher than normal furnace temperatures. Wet bottom furnaces are thus not suitable.

4.1 Dry Bottom Suspension Combustion

By elimination, therefore, dry bottom pulverized fuel, combustors are believed to be the relatively risk-free type for all systems. This equipment can easily achieve the necessary efficiencies and capacities; it is known to be highly reliable and corrosion resistant. Horizontal firing is the design choice for the lower S/C ratio fuels and vertical down-firing for high S/C ratio fuels as discussed in Part XI.

5. COMBUSTION PROCESS DESIGN

Figure IV-2 is a generalized diagram of air and combustion gas distribution that is typical of suspension firing of pulverized fuels. A similar gas flow is employed in the combustors of CASE A, C and D. The combustor of CASE D is suitable for CASE E with minor changes. The CASE B combustor design has not been developed for this study since funds were not provided for it, but an estimate of capital cost based on a modified CASE A design is indicated in Part XIV.

Table IV-6 is a theoretical representation of expected change in Boiler Performance with change in FeS_2 concentration in the HSC Fuels. Note that decline in efficiency across the entire range is not great. This relation is plotted in Figure IV-3. The chart shows also that flue gas quantity and heating value increase at about the same rate as the coal content in the pyrite-coal mixture. Flue gas volume for a given heat output is thus independent of the sulfur content of the fuel, and the SO_2 concentration in the flue gas will be substantially controlled by the S/C ratio as shown in Figure III-1.

Table IV-7 and the diagrams following are a summary of the Boiler Specifications appearing in Part XI. Steam quality specifications are established for typical industrial applications, or as such specifications are usually laid down by manufacturers of turbo-generators for driving with sub-critical steam.

FIGURE IV-2

COMBUSTION PROCESS
FLOW SHEET
DISTRIBUTION OF PREHEATED AIR

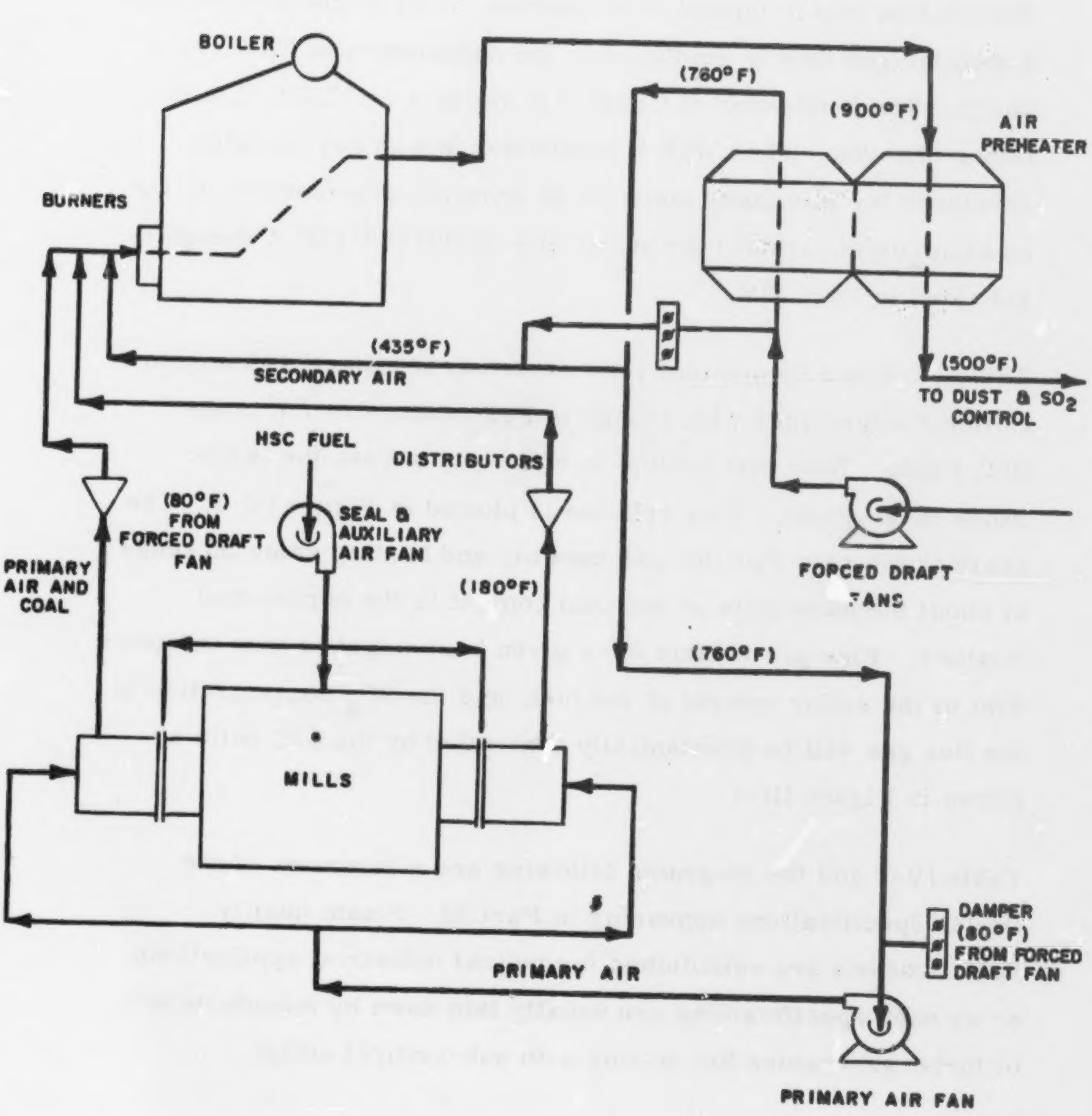

BOILER
BURNERS
(760°F)
(900°F)
AIR PREHEATER
(435°F)
SECONDARY AIR
(500°F)
TO DUST & SO_2 CONTROL
HSC FUEL
DISTRIBUTORS
(80°F) FROM FORCED DRAFT FAN
SEAL & AUXILIARY AIR FAN
PRIMARY AIR AND COAL
(180°F)
FORCED DRAFT FANS
(760°F)
MILLS
DAMPER (80°F) FROM FORCED DRAFT FAN
PRIMARY AIR
PRIMARY AIR FAN

TABLE IV-6

EXPECTED HSC FUEL-FIRED COMBUSTOR PERFORMANCE

Composition of Fuel - Combustibles Only:					
% FeS_2	0	25	50	75	100
% Coal	100	75	50	25	0
S/C Ratio	0	0.18	0.53	1.6	∞
Heat Losses, %					
Dry Flue Gas	9.9	10.2	10.7	11.8	14.5
Hydrogen + Water in the fuel	4.3	4.2	4.0	3.6	2.5
Moisture in the Air	.23	.23	.24	.26	.32
Unburned Carbon	.1	.31	.66	1.29	3.2
Radiation	.38	.38	.38	.38	.38
Unaccounted for	1.5	1.5	1.5	1.5	1.5
Total	16.4	16.8	17.5	18.8	22.4
Boiler Efficiency %	83.6	83.1	82.5	81.2	77.6
Flue Gas/Fuel Weight Ratio	13.7	11.4	9.0	6.7	4.3
Air/Fuel Weight Ratio	12.7	10.5	8.3	6.1	3.9
MBTU per POUND ca.	14	11.3	8.5	5.8	3.0

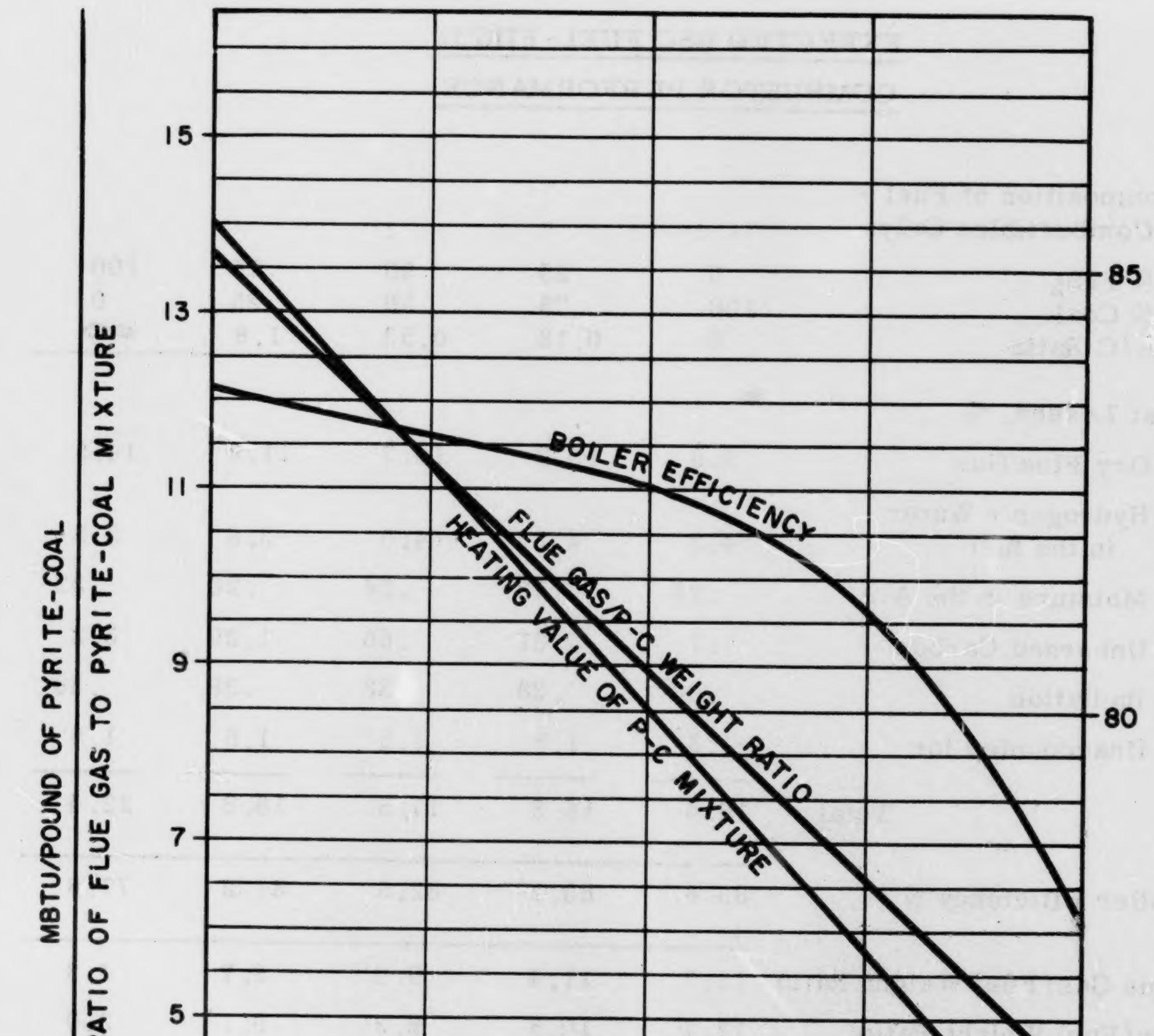
Moisture in actual HSC fuel 6%.

Non-pyritic ash in actual HSC Fuel 20%.

Combustion in 25% excess air.

Humidity of air 60% RH at 80° F.

FIGURE IV-3



COMBUSTION EFFECTS
OF
PYRITE-COAL MIXTURES

*WITH 25% EXCESS AIR

SUMMARY OF HIGH SULFUR COMBUSTION PROCESSES

	Unit	CASE A	CASE B	CASE C	CASE D	CASE D_1	D_2	CASE E
						500 MW Power		
		Industrial Steam and Sulfuric Acid	Industrial Steam and Sulfuric Acid (Monsanto Process)	Prototype 50 MW Power & Sulfur	500 MW Power & Sulfur	Magnesium Sulfite	Liquid Sulfur Dioxide	500 MW Power and Sulfuric Acid (Monsanto Process)
HSC FUEL:	Sulfur/Coal	1.8	0.4	0.2	0.2	0.2	0.2	0.12
FeS_2	% Dry Basis	61.6	33.8	20.8	20.8			13.2
Organic S	" " "	0.2	0.5	0.6	0.6			0.9
Total S	" " "	(33.1)	(18.5)	(11.7)	(11.7)			(7.9)
Coal Combustibles	" " "	18.2	45.7	58.6	58.6			65.9
Non-Pyritic Ash	" " "	20.0	20.0	20.0	20.0			20.0
Moisture	% Wet Basis	6.0	6.0	6.0	6.0			6.0
Heating Value	Btu/Lb	4424	7400	9000	9000			9660
COMBUSTOR			NOTE (1)					NOTE (1)
Fuel Flow	MLbs/Hr	160	101.4	74.4	570			533
Gross Heat Input	10^6 Btu/Hr	700	750	670	5128			5150
Net Heat Output	10^6 Btu/Hr	569	569	551.4	4220			4220
Boiler:Steam	MLbs/Hr	500	500	500	3500			3500
Temp./Press.	°F/Psig	825/900	825/900	1000/1450	1005/2500			1005/2500
Generator:Electricity	MW			50	500	500	500	500
Gross Heat Rate	Btu/Kwh			13,400	10,250			
TOTAL ASH	Tons/Day	1171	517	303	2315			1842
Pyritic Ash	Tons/Day	787	274	124	947			562
Iron	Tons/Day	551	192	86	662			393
FLUE GAS	MSCFM	177	173	167	1110			1157
Particulate Load	Grains/SCF	50	40	17	17			
SO_2	Vol. % Dry	6.0	2.0	1.15	1.15			0.7
SO_3	" " "	0.2	0.1	0.046	0.046			0.02
CO_2	" " "	8.6	14.0	15.4	15.4			15.9
O_2	" " "	4.3	4.2	4.26	4.26			4.2
N_2	" " "	80.9	79.7	79.1	79.1			79.18
H_2O Nominal	Vol. % Wet	4	4	4	4			4
Temperature	°F	500	805	500	500			850
SULFUR VALUE IN FLUE GAS								
Sulfur	Tons/Day			102	770			
Sulfur as H_2SO_4	" "	1831	624					1422
Sulfur as $MgSO_3$	" "					2500		
Sulfur as Liq. SO_2	" "						1520	

NOTE (1) Combustor data for Case B and E are reported tentatively pending detailed analysis of these conditions.

(2) All "Tons" are short or net tons.

TABLE IV-7

6. INDUSTRIAL BOILER

The Industrial Boiler of 500, 000 pounds per hour steam capacity, when designed for suspension firing and dry ash, is adaptable to HSC Fuels in the range S/C = 1 to S/C = 2. 5 provided the non-pyritic ash is not excessive.

6. 1 Vertical Firing

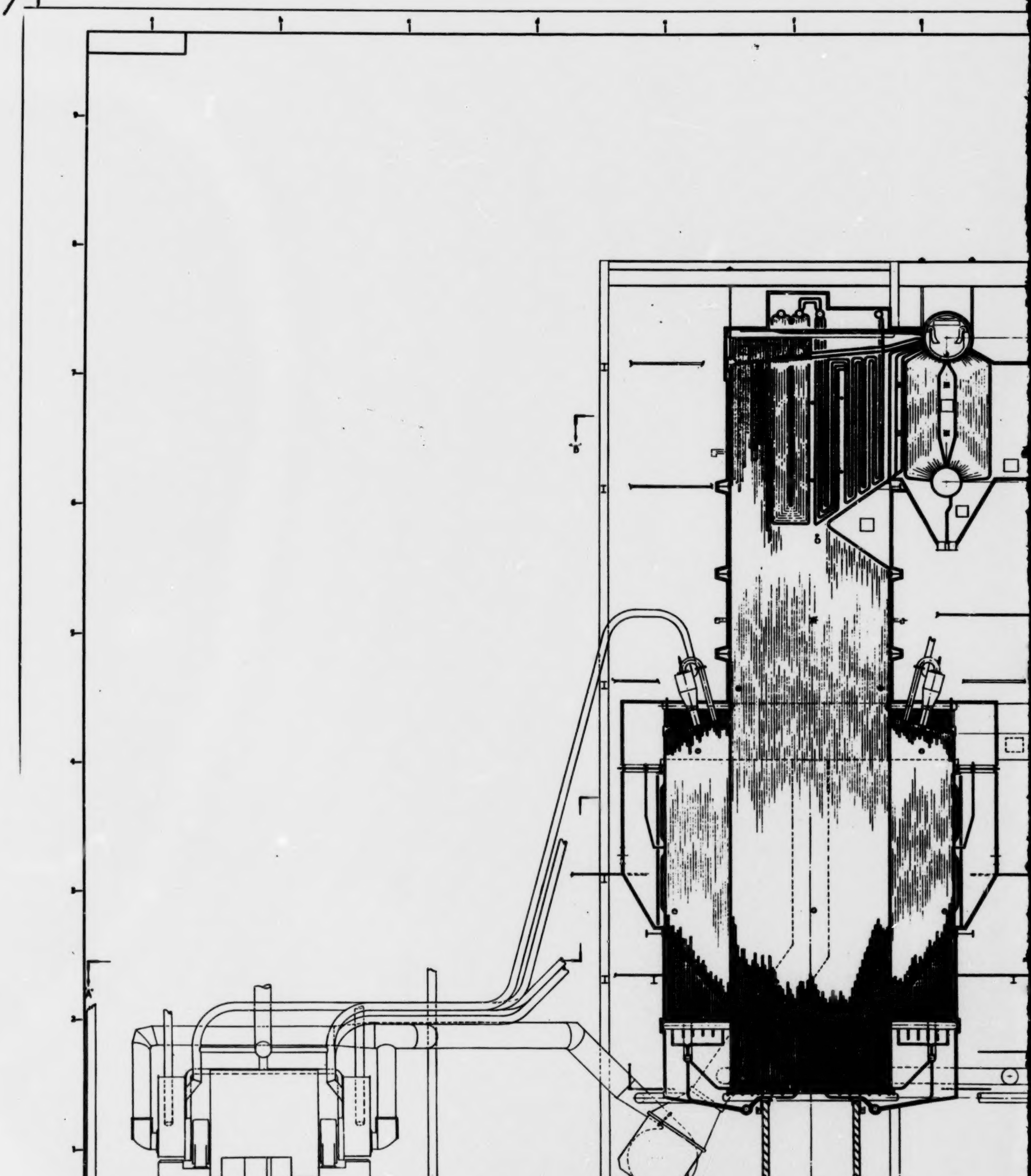
The high sulfur, high total ash, slower combustion and slightly reduced volatility of fuels in this S/C range, as compared to commercial steam coals, dictate a preference for special burners, Figure IV-5, fired vertically downward, as illustrated in the boiler elevation drawing of Figure IV-4. This arrangement permits a long flame path and allows introduction of air along the path of combustion as needed in the event of variation in the combustion rate of the HSC Fuel. A portion of the preheated primary air is diverted through the pulverizing mills to convey the fuel as shown in Figure IV-2. Ball tube mills, a preferred type for heavy duty, appear to be required for all the HSC Fuels in view of the low grindability of iron pyrite. (See Part XI for elaboration.)

6. 2 Steam cycle design at 500, 000 Lbs/Hr output flow is:

Pressure	900 psig
Temperature	825° F
Reheat	None
Feedwater	300° F

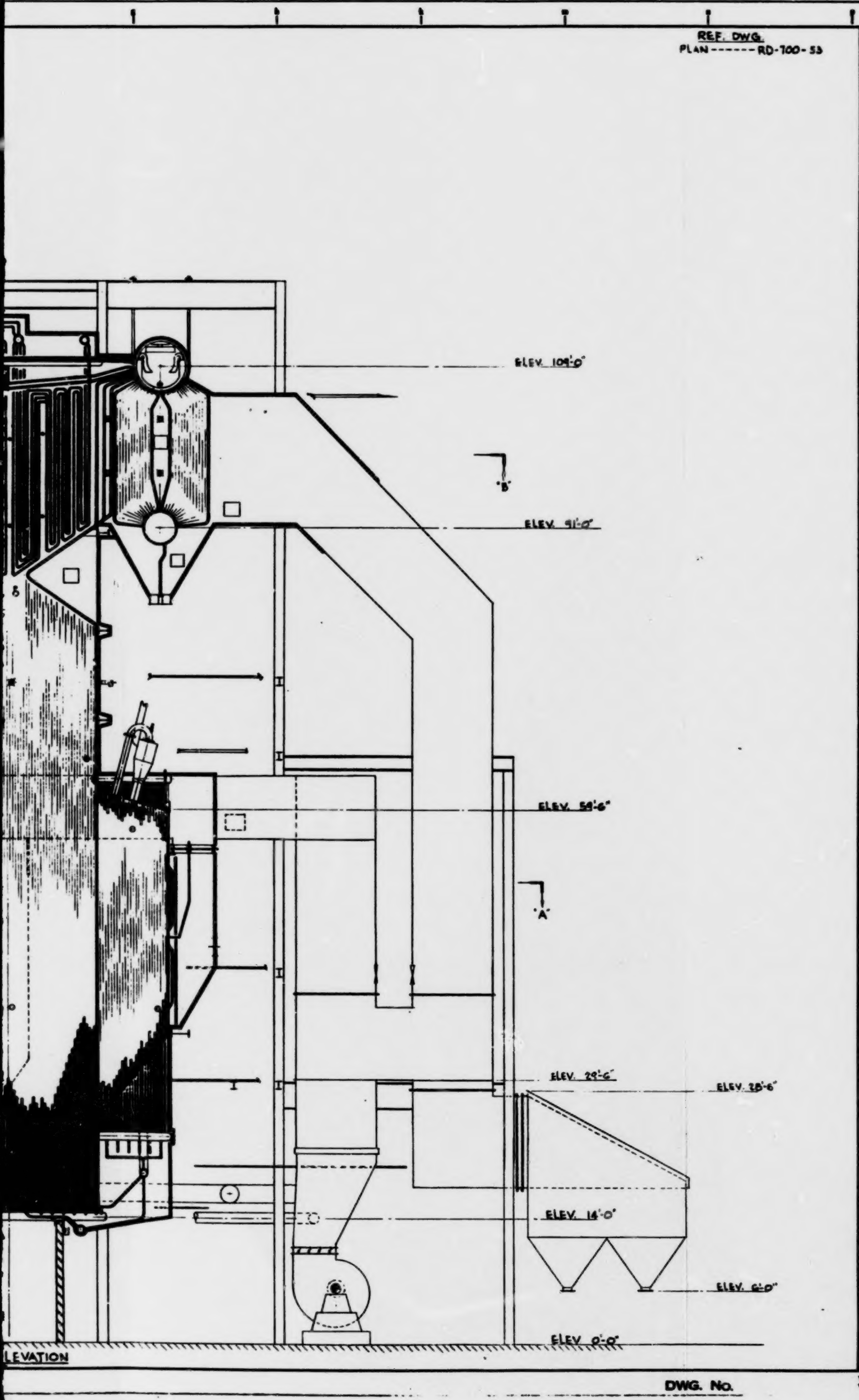
The gas temperature is 500° F at the air heater exit. This level is safely above the dew point of condensable acids in the air heater but in this unit the economizer has been eliminated in view of the exceptionally high acidic concentration in the flue gas.

A



SECTIONAL SIDE ELEVATION

B


REF. DWG.
PLAN ------ RD-700-53
ELEV. 109'-0"
'B'
ELEV. 91'-0"
ELEV. 54'-6"
'A'
ELEV. 29'-6"
ELEV. 28'-6"
ELEV. 14'-0"
ELEV. 6'-0"
ELEV. 0'-0"
LEVATION
DWG. No.

REVISIONS

500,000 LB/HR
INDUSTRIAL
HIGH SULFUR COMBUSTOR
(ELEVATIONAL VIEW)

3/32"

RD-700-52

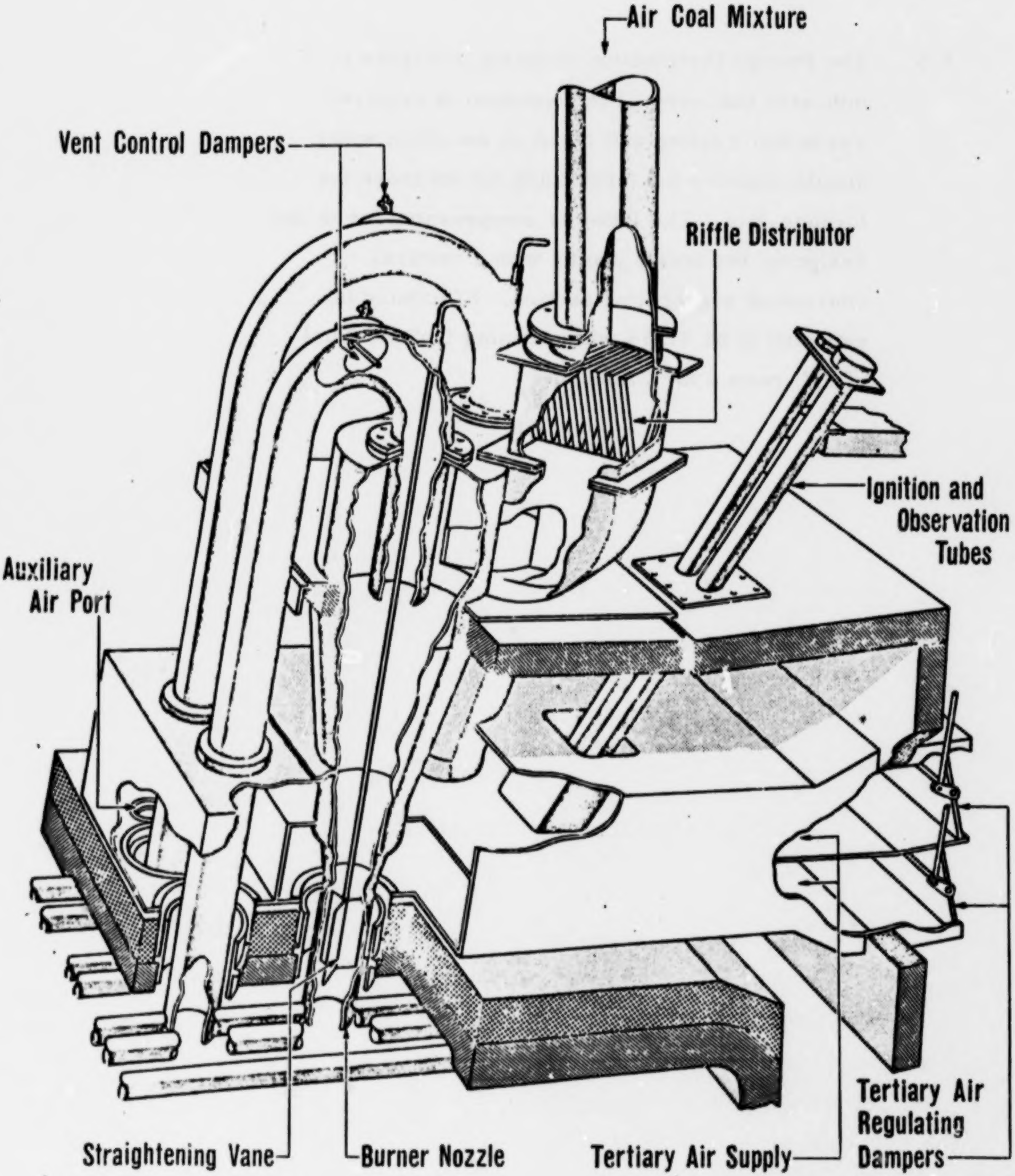
0-59-1177

T. S. 5/19/70

FIGURE IV - 4

FIGURE IV-5

'CYCLONE' ANTHRACITE BURNER

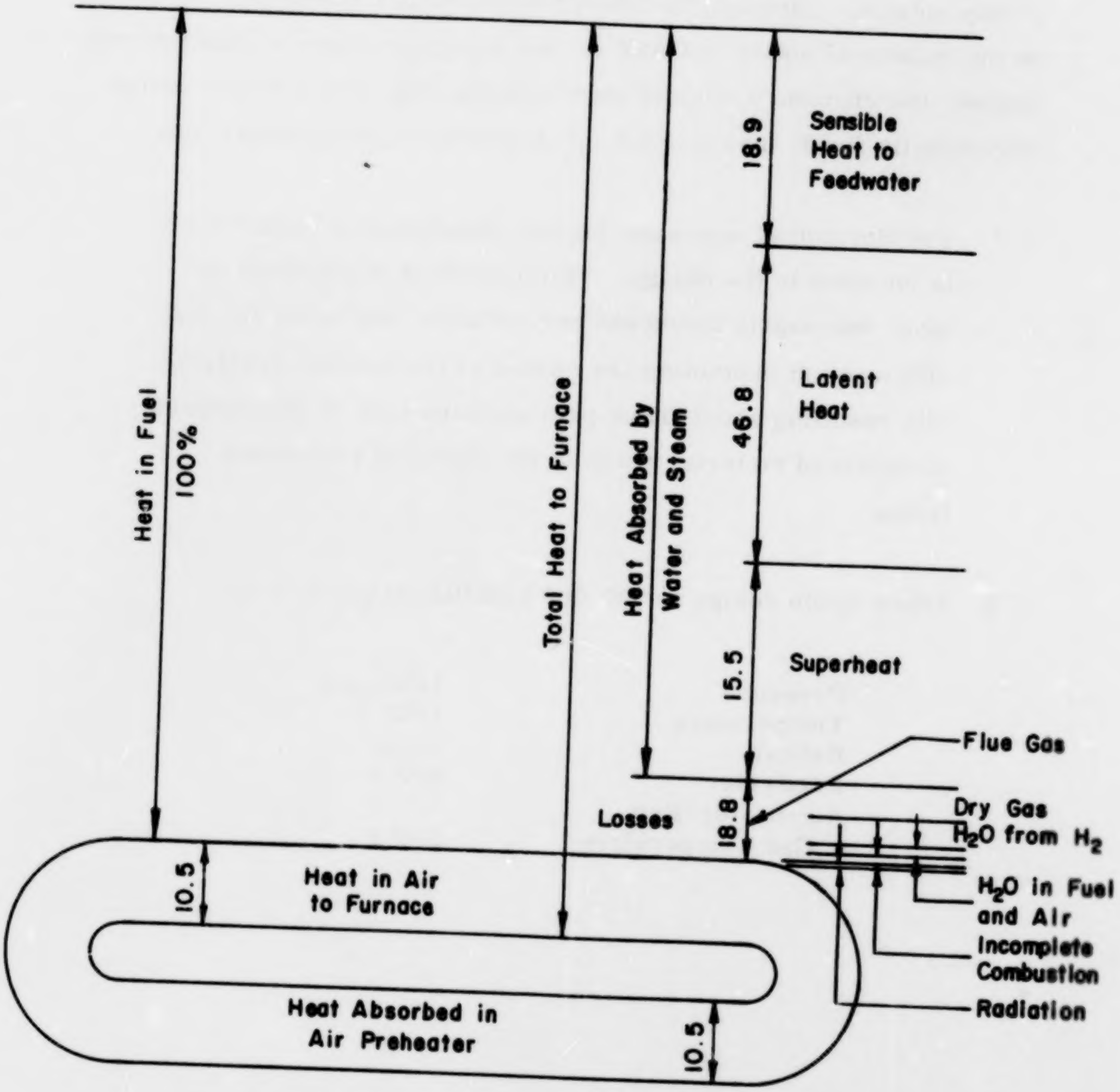

Air Coal Mixture
Vent Control Dampers
Riffle Distributor
Ignition and Observation Tubes
Auxiliary Air Port
Straightening Vane
Burner Nozzle
Tertiary Air Supply
Tertiary Air Regulating Dampers

6.3 The Energy Distribution Diagram of Figure IV-6 indicates that very little superheat is required. Feedwater heating will occur in the drum which should improve the circulation but increase the furnace duty. The latter is compensated for in the design by two boiler drums with a vertical, convective evaporating section. Efficiency is expected to be 81.3% when burning the HSC Fuel of S/C ratio 2 for CASE A.

FIGURE IV - 6

DISTRIBUTION OF ENERGY
500,000 LB/HR BOILER
FOR CASE A

HSC FUEL S/C = 2


Heat in Fuel
100%
Total Heat to Furnace
Heat Absorbed by Water and Steam
18.9
Sensible Heat to Feedwater
46.8
Latent Heat
15.5
Superheat
Flue Gas
Losses
18.8
Dry Gas
H_2O from H_2
H_2O in Fuel and Air
Incomplete Combustion
Radiation
10.5
Heat in Air to Furnace
Heat Absorbed in Air Preheater
10.5

7. THE 50 MEGAWATT PROTOTYPE BOILER FOR CASE C

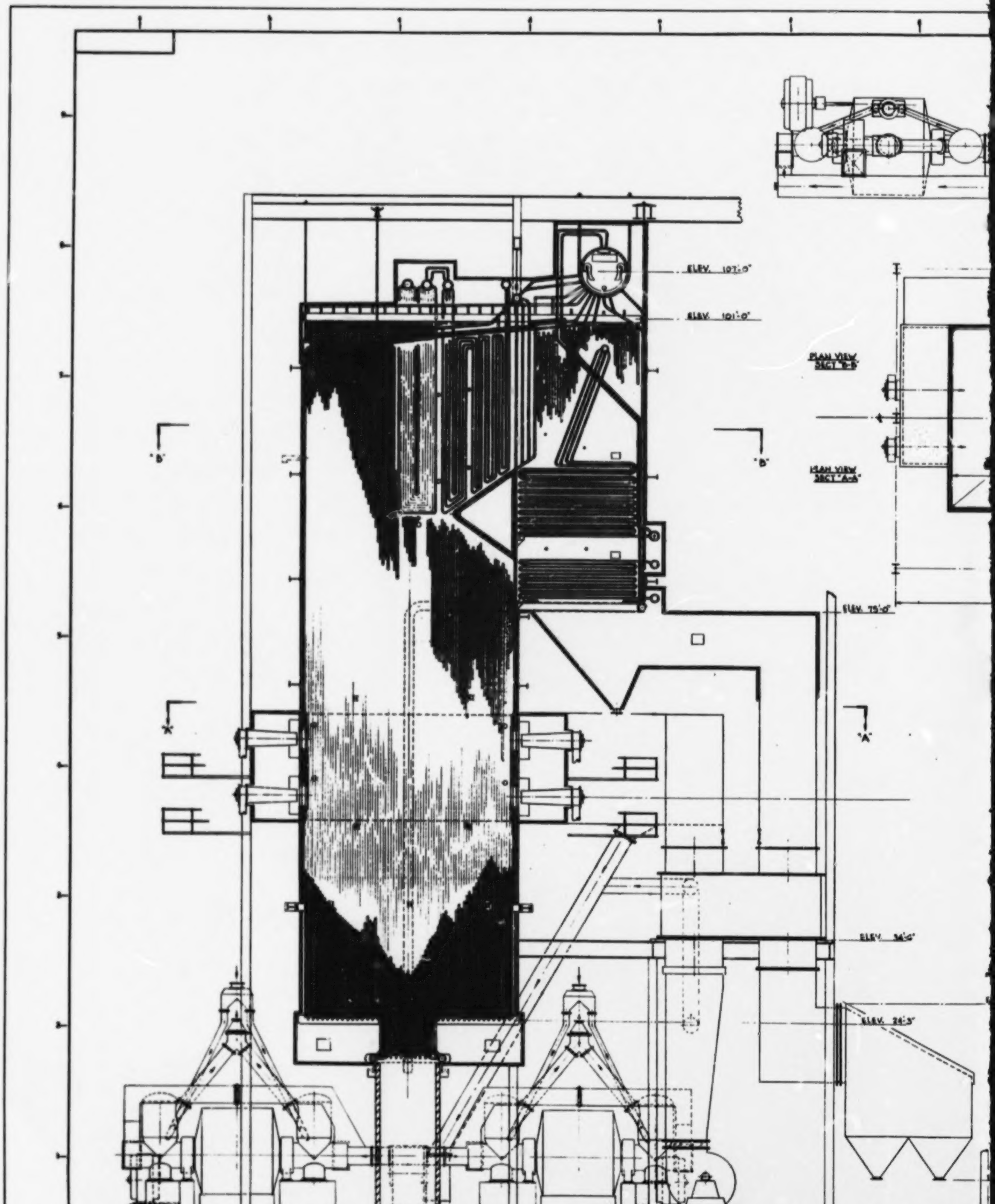
All designs of the CASE C system are consistent with the conception of an electric utility demonstration plant as described in Part I. 2. 6 of this volume. Although the energy output is of a magnitude similar to the industrial boiler of CASE A, the steam pressure is substantially higher, the efficiency slightly improved (82. 3%), and a boiler design, horizontally fired, that is much more nearly of conventional type.

7. 1 The Horizontal Intervane Burner illustrated in Figure IV-8 is included in the design. This burner is of conventional type, thoroughly developed and reliable, and noted for high efficiency in promoting the mixing of combustion reactants. The resulting short flame path obviates risk of impingement of unburned material which is the object of suspension firing.

7. 2 Steam cycle design at 500, 000 LBS/HR output flow is:

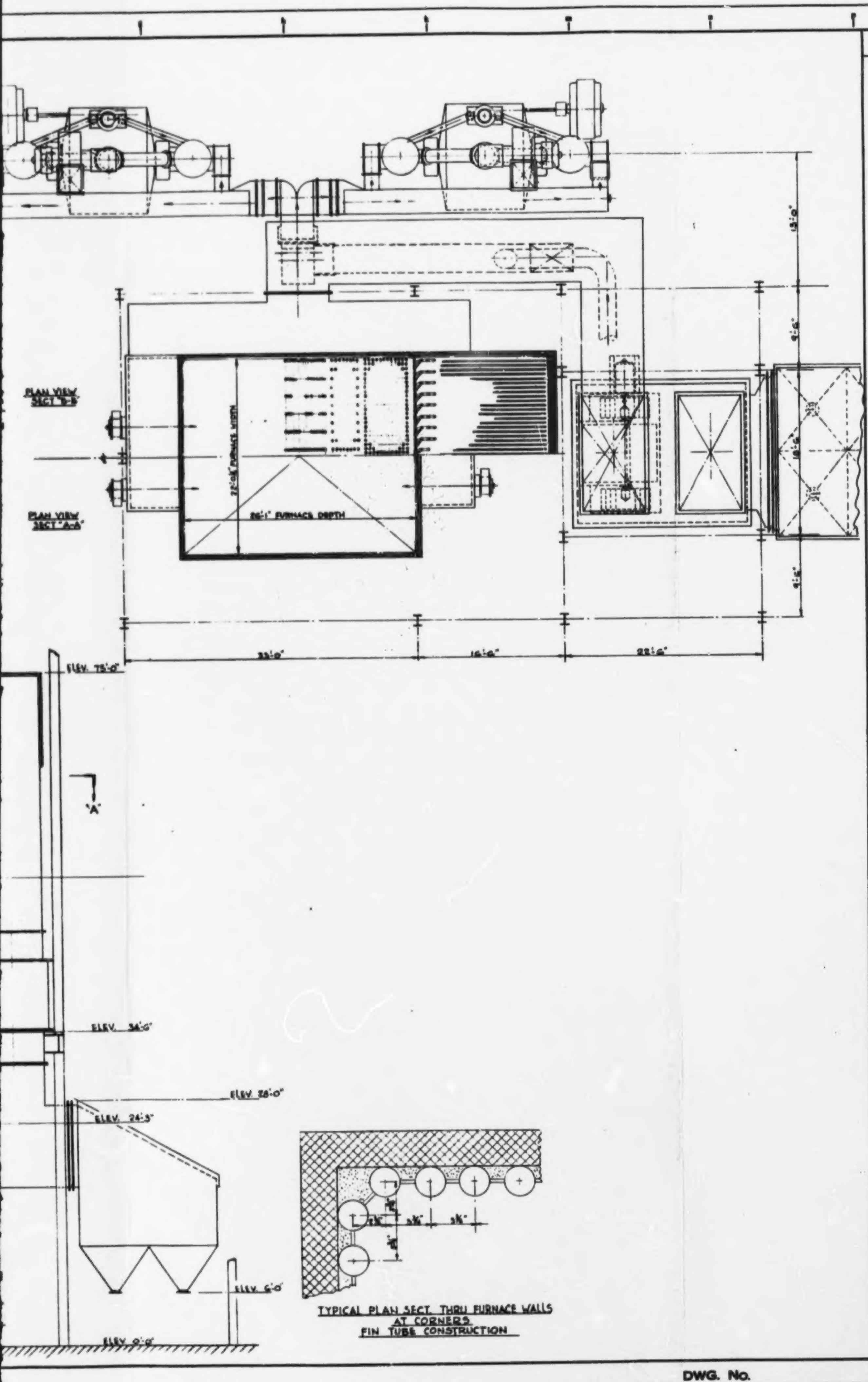
Pressure	1450 psig
Temperature	1000° F
Reheat	None
Feedwater	430° F
Air-heater-Exit Gas Temperature	500° F

a


ELEV. 107'-0"
ELEV. 101'-0"
PLAN VIEW SECT "B-B"
"B"
"B"
PLAN VIEW SECT "A-A"
ELEV. 75'-0"
"A"
"A"
ELEV. 34'-0"
ELEV. 24'-3"
ELEV. 0'-0"

SECTIONAL SIDE ELEVATION

B


PLAN VIEW SECT "B-B"
PLAN VIEW SECT "A-A"
FURNACE WIDTH
26'-1" FURNACE DEPTH
13'-0"
9'-6"
18'-6"
9'-6"
33'-0"
16'-0"
22'-6"
ELEV. 75'-0"
'A'
ELEV. 34'-0"
ELEV. 28'-0"
ELEV. 24'-3"
ELEV. 6'-0"
ELEV. 0'-0"
DWG. No.

TYPICAL PLAN SECT. THRU FURNACE WALLS
AT CORNERS
FIN TUBE CONSTRUCTION

NOTES

REVISIONS

50 MW
PROTOTYPE
HIGH SULFUR COMBUSTOR

RD-700-51

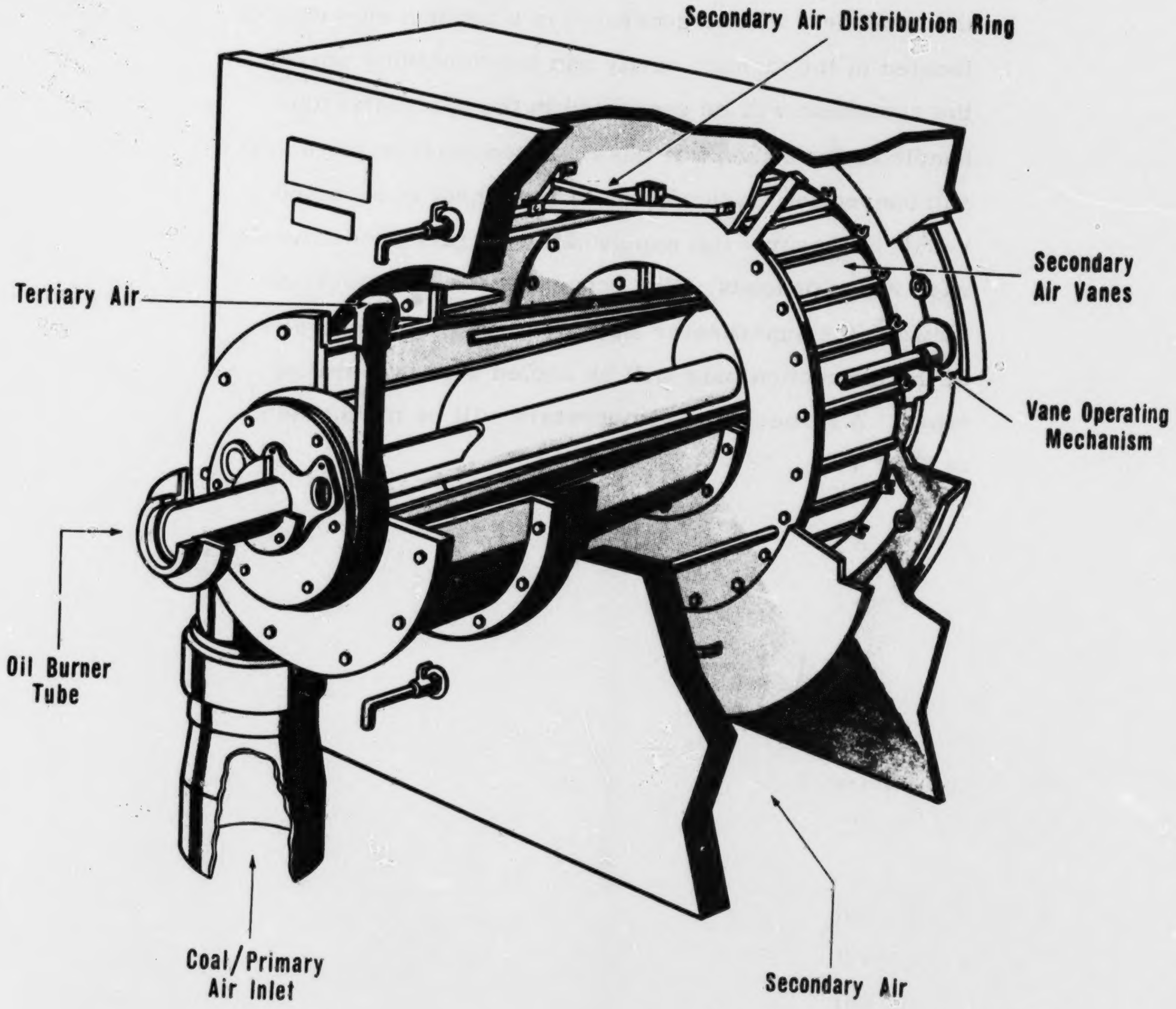
0-59-1177

T.S. 5/4/70

FIGURE IV - 7

FIGURE IV-8

INTERVANE BURNER

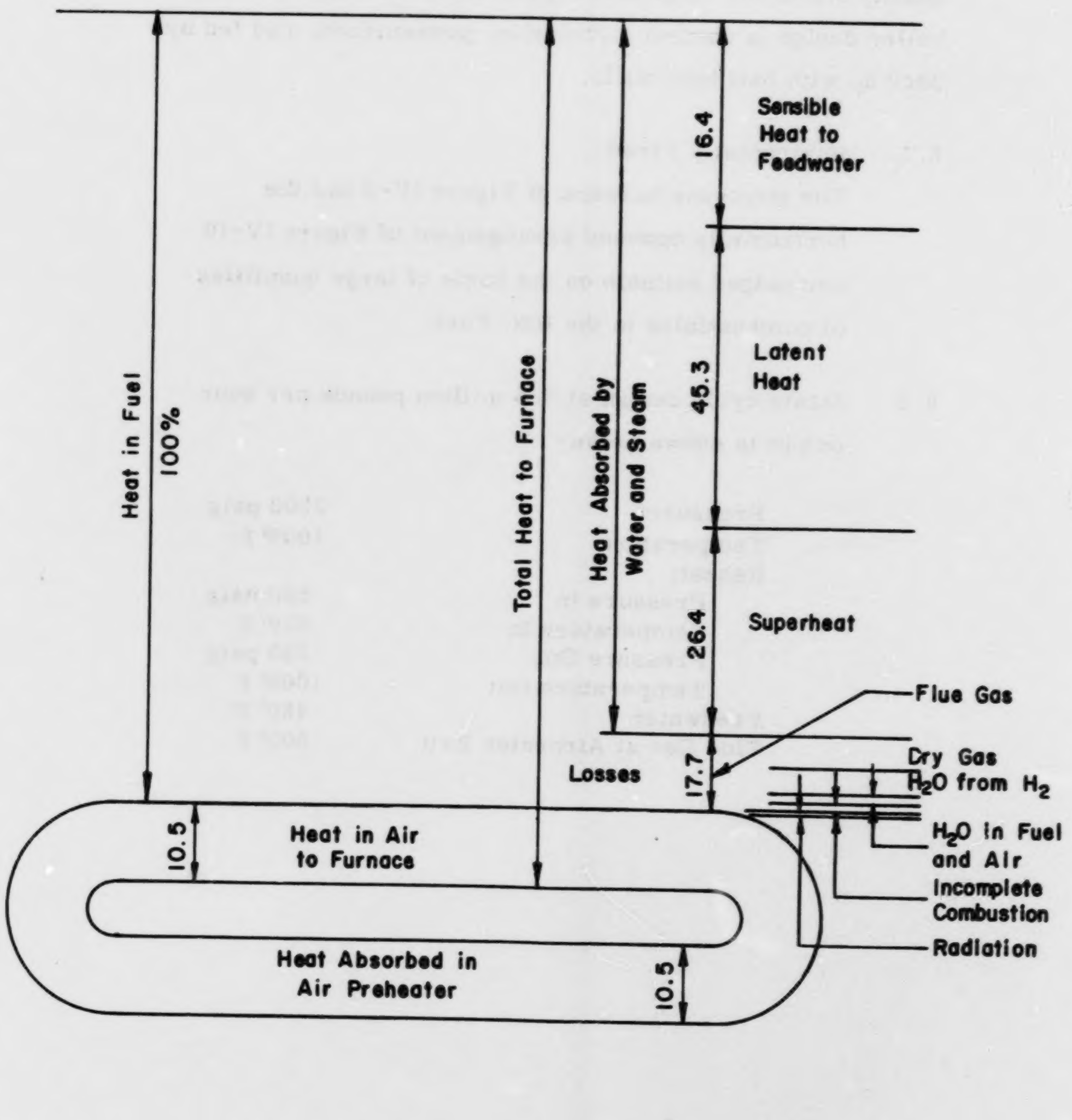

Secondary Air Distribution Ring
Secondary Air Vanes
Tertiary Air
Vane Operating Mechanism
Oil Burner Tube
Coal/Primary Air Inlet
Secondary Air

7.3 The Energy Distribution diagram of Figure IV-9 is quite different from that of the Industrial Boiler. A portion of the superheat will be generated in a pendent superheater located in the furnace cavity and the remaining portion of the superheat will be generated in the horizontal tube bundle in the convection pass. A combination of radiant and convection transmission of superheat is expected to result in a nearly flat superheater characteristic over a wide range of loads. The economizer will be located between the superheater and the air heater. The walls of the convection pass will be cooled with evaporating tubes. A furnace exit temperature will be maintained at 2000° F.

FIGURE IV-9

DISTRIBUTION OF ENERGY
50 MEGAWATT BOILER
FOR CASE C

HSC FUEL S/C = 0.2


Heat in Fuel
100%
Total Heat to Furnace
Heat Absorbed by Water and Steam
16.4
Sensible Heat to Feedwater
45.3
Latent Heat
26.4
Superheat
Flue Gas
17.7
Losses
Dry Gas
H_2O from H_2
H_2O in Fuel and Air
Incomplete Combustion
Radiation
10.5
Heat in Air to Furnace
Heat Absorbed in Air Preheater
10.5

8. THE 500 MEGAWATT ELECTRIC UTILITY BOILER

While the boiler selected for this duty is to burn the same HSC Fuel, S/C = 0.2, specified for the prototype, CASE C, higher quality steam for large turbo-generators is required. The boiler design is natural circulation, pressurized, and fed by back up with ball tube mills.

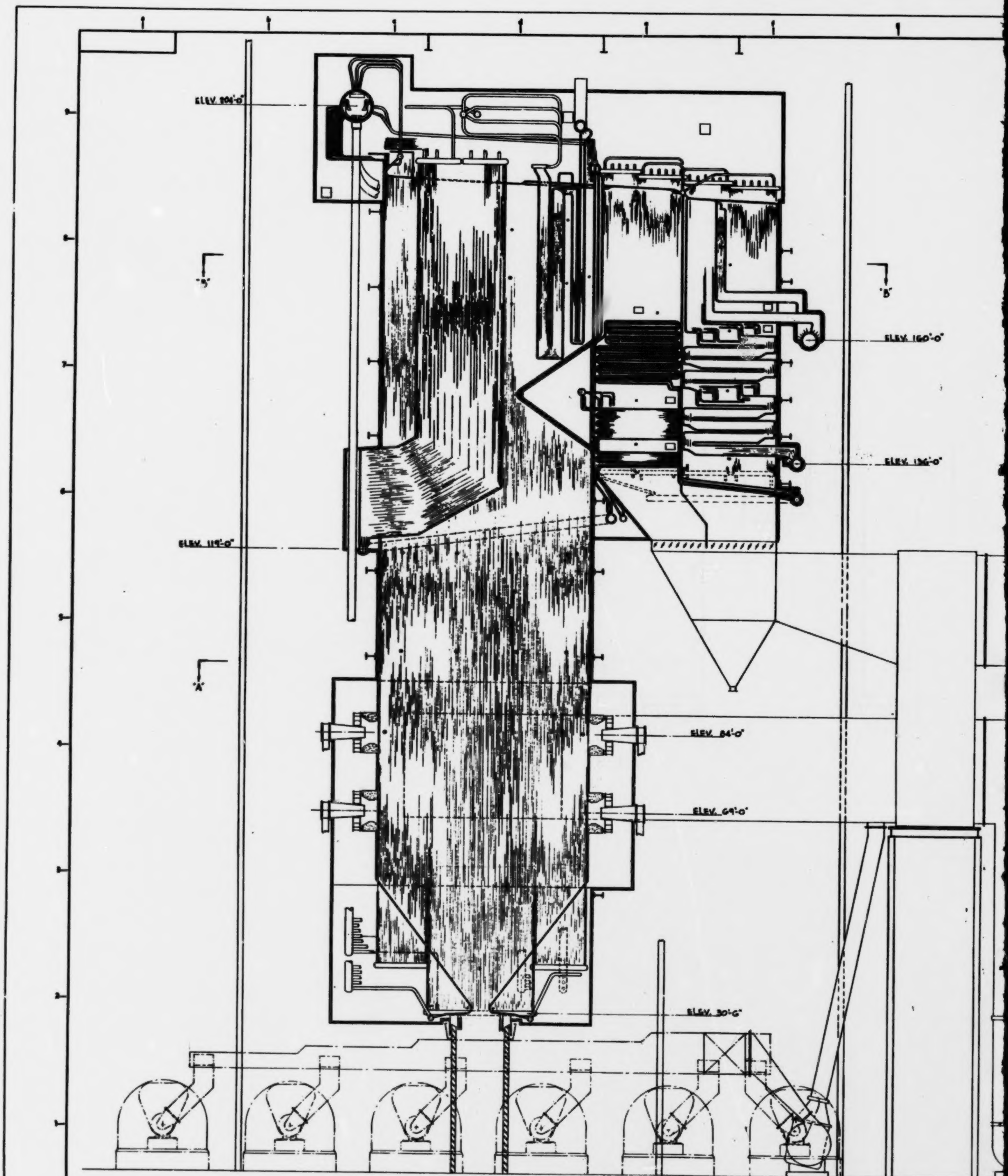
8.1 Horizontally Fired

The intervane burners of Figure IV-8 and the horizontally opposed arrangement of Figure IV-10 are judged suitable on the basis of large quantities of combustibles in the HSC Fuel.

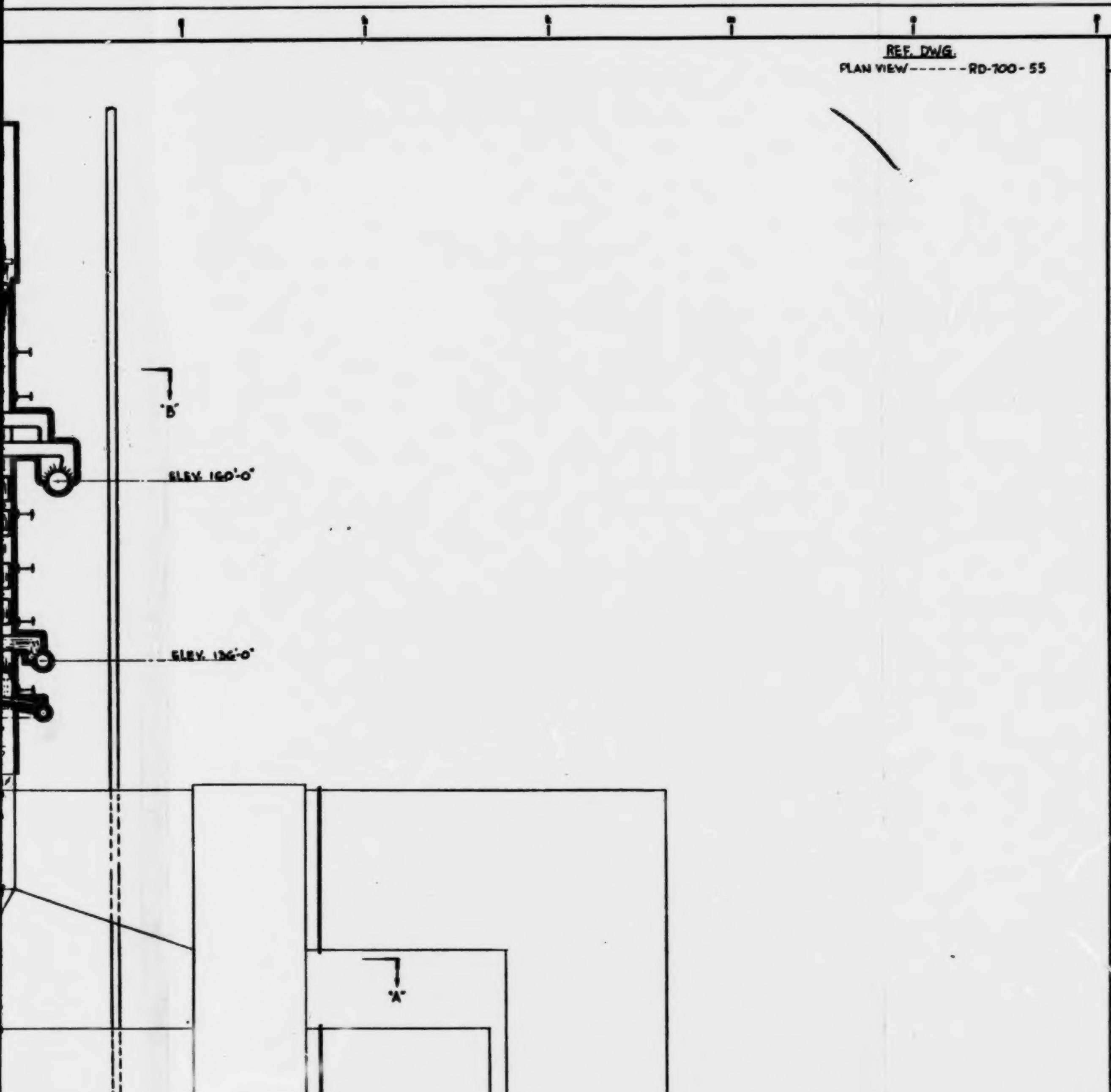
8.2 Steam cycle design at 3.5 million pounds per hour output is shown below:

Pressure	2500 psig
Temperature	1005° F
Reheat:	
Pressure In	580 psig
Temperature In	629° F
Pressure Out	555 psig
Temperature Out	1005° F
Feedwater	480° F
Flue Gas at Airheater Exit	500° F

4


ELEV. 204'-0"
"B"
"B"
ELEV. 160'-0"
ELEV. 136'-0"
ELEV. 119'-0"
"A"
ELEV. 84'-0"
ELEV. 69'-0"
ELEV. 30'-6"

B


REF. DWG.
PLAN VIEW------RD-700-55
"B"
ELEV. 160'-0"
ELEV. 136'-0"
"A"
ELEV. 33'-0"
ELEV. 6'-0"
ELEV. 0'-0"
DWG. No.

500 MW

HIGH SULFUR COMBUSTOR
(ELEVATIONAL VIEW)

RD-700-54

0-59-1177

T.S. 6-8-70

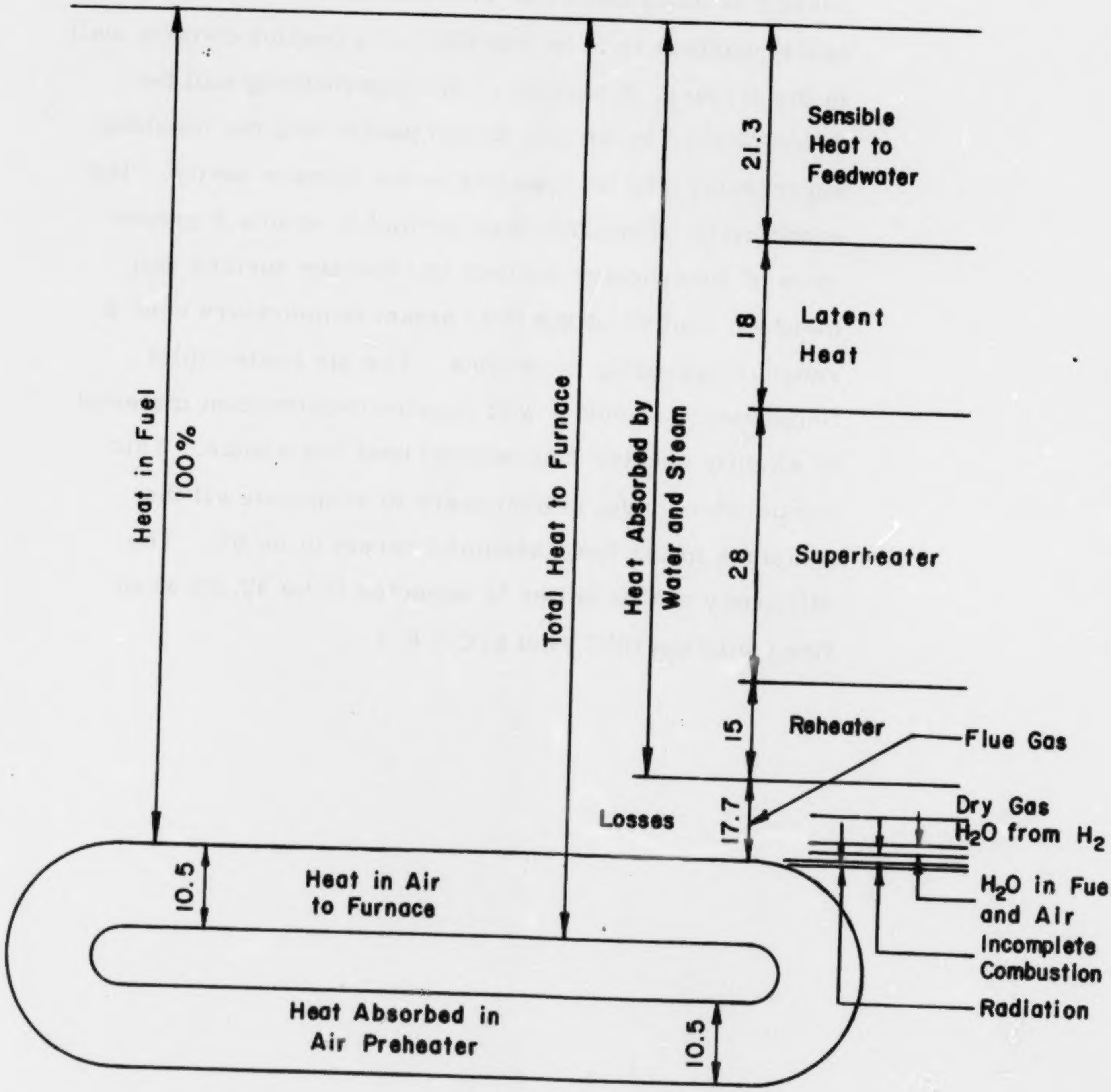
FIGURE IV - 10

8.3 The Energy Distribution diagram of Figure IV-11 shows that evaporation accounts for a small portion only of the total boiler duty. In order to minimize startup problems, overheating of tubes and other undesirable effects, the superheater surface is to be installed as a partial division wall in the furnace. A portion of the superheating will be accomplished in the convection passes and the finishing superheater will be installed in the furnace cavity. The economizer is smaller than normal to enable a proper ratio of superheater surface to reheater surface and maintain control of the final steam temperature over a range of operating conditions. The air heater inlet temperature at 900° F will require construction material of slightly greater than normal heat resistance. This temperature level is necessary to evaporate all the moisture in the fuel, assumed herein to be 6%. The efficiency of this boiler is expected to be 82.3% when fired with the HSC Fuel S/C = 0.2.

FIGURE IV-11

DISTRIBUTION OF ENERGY
500 MW BOILER
FOR CASE D

HSC FUEL S/C = 0.2


Heat in Fuel
100%
Total Heat to Furnace
Heat Absorbed by Water and Steam
21.3
Sensible Heat to Feedwater
18
Latent Heat
28
Superheater
15
Reheater
Flue Gas
17.7
Losses
Dry Gas
H_2O from H_2
H_2O in Fuel and Air
Incomplete Combustion
Radiation
10.5
Heat in Air to Furnace
Heat Absorbed in Air Preheater
10.5

9. BOTTOM ASH AND FLY ASH

Bottom ash and fly ash are to be collected mechanically and by wet scrubbing from the boilers of CASE A, C and D Systems in about the following proportions of the total ash:

Furnace Bottom	20%
Cyclone Collector	65%
Wet Scrubber	15%
Total	100%

The dry cyclones in series with wet scrubbing is well justified by reduction in the volume of ash that will need to be processed through lime neutralization of the acidity of the scrubbing liquor.

Fly ash is to be collected by electric precipitation in CASE B and E Systems. The design basis for fly ash collection from all systems is set forth in Table IV-8.

9.1 While the iron content of the dry CASE A ash is as high as 47% it is not likely to be a valuable source of blast furnace charge. The other ashes are lower in iron. Possibilities for development of economic value from the ashes is discussed in Part XII.

TABLE IV-8

FLY-ASH COLLECTION DESIGN BASIS

Case		A	C	D
Capacity		500,000 lb/hr	50 MW	500 MW
Gas Flow lb/hr		895,000	675,000	5,170,000
Gas Temperature °F		496	502	497
ΔP			About 3" H_20	
Particle Loading				
Total Ash, Grains/SCF		50	17	17
To Dust Collector				
Grains/SCF		40	13.5	13.5
Ash Composition+				
Fe_20_3	Weight Percent	80	80	80
Al_20_3 $Si0_2$		17	17	17
Remaining		3	3	3
Efficiency Required			75 - 80 %	
Ash Size Distribution		85%*	45µ	
		80%	40µ	
		65%	20µ	
		40%	10µ	
		23%	5µ	

+ Nominal

* Percent under indicated micron size

V

ELECTRIC GENERATION

1. SIZING OF STEAM-ELECTRIC EQUIPMENT

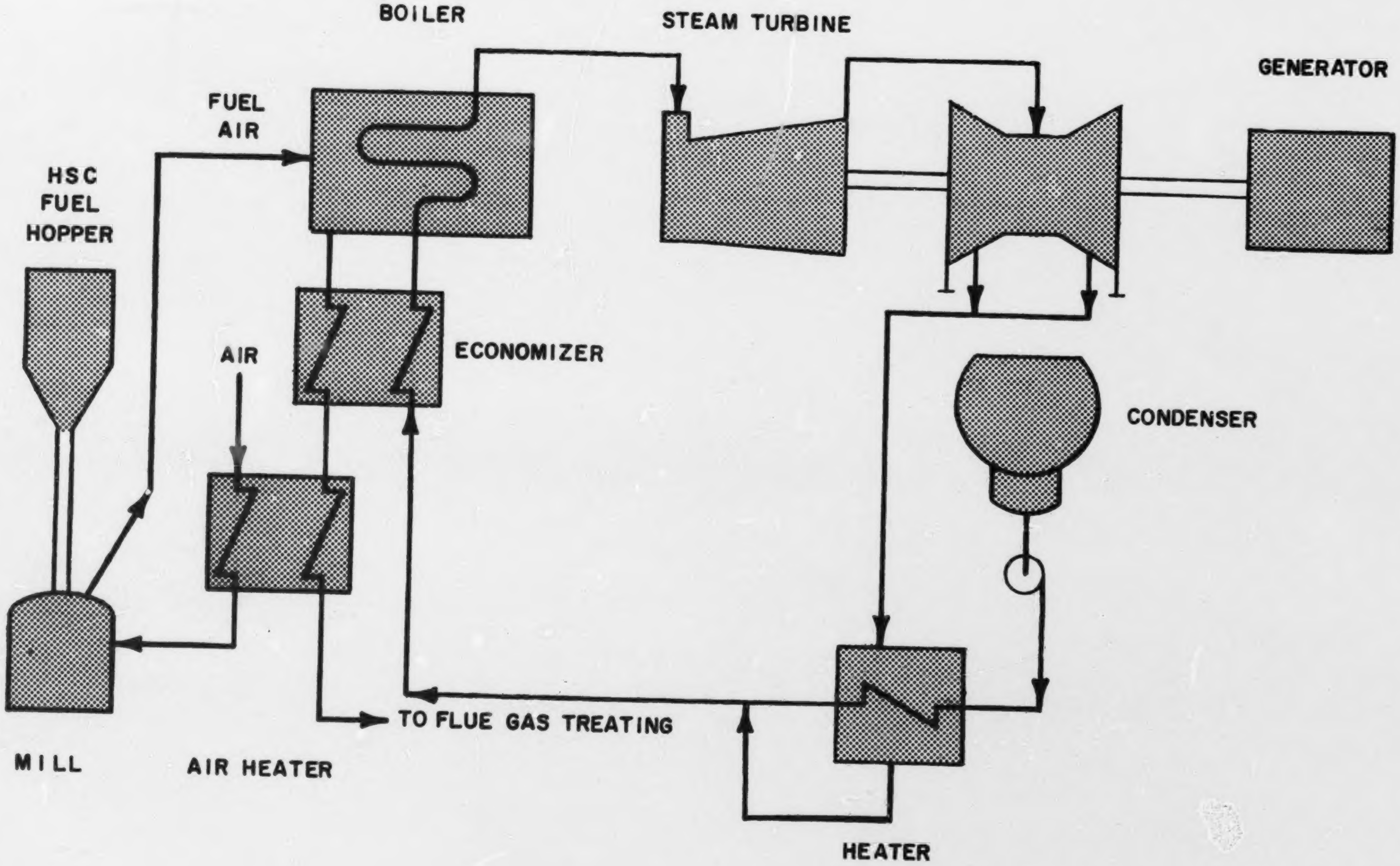
The 500 MW unit is among the smaller of the popular sizes of generating units purchased in very recent years, most of which have ranged from 600 to 850 MW capacity per unit. Unit sizes of 300 to 500 MW were typical of new U.S. installations built during the 7 or 8 years prior to 1967. These sizes were perhaps a little better than the larger units. In any event they enjoy a reputation for somewhat greater reliability. The choice of the 500 MW unit, (specified in the statement of contract work) is a choice for conservative technology which is a good companion to available technology also emphasized in the statement of contract work. For much the same reasons the 500 MW boiler is designed for subcritical operation and is otherwise consistent with conservative technology.

The 50 MW boiler is also sub-critical and conservative and is of a size now and then installed for municipalities, industrial service and the like. While the steam pressure is much lower than is common for boilers 10 to 20 times larger, it will be completely adequate to demonstrate the technical feasibility of converting the heat value in the HSC Fuel to electricity in large scale equipment. Steam-electric generators much smaller than 50 MW tend to be classified by manufacturers as "special order" with a higher dollars per kilowatt price tag. The 50 MW size thus appeared about right for the prototype plant.

2. **A conventional turbine drive cycle is illustrated in Figure V-1 appropriate for both 50 and 500 MW units.**

3. **Central Station Steam Generators as installed in recent years are listed in Table V-1. All units in the list are 100 MW or larger.**

FIGURE V-1


BOILER
STEAM TURBINE
GENERATOR
FUEL
AIR
HSC
FUEL
HOPPER
AIR
ECONOMIZER
CONDENSER
MILL
AIR HEATER
TO FLUE GAS TREATING
HEATER

PULVERIZED HSC FUEL/STEAM/POWER PLANT

TABLE V-I

CENTRAL STATION
STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulvs. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Public Service Co. Of Indiana, Inc. Plainfield, Indiana Wabash River Generating Station West Terre Haute, Indiana Unit #1	700,000 100,000	1700 1500	1005 1005	Pulv. Coal	2-D7 EX	Bal. Draft	Nat Circ.	Dry Ash	No	9/50	6/53
Public Service Co. Of Indiana, Inc. Plainfield, Indiana Wabash River Generating Station West Terre Haute, Indiana Unit #2	700,000 100,000	1700 1500	1005 1005	Pulv. Coal	2-D7 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	9/50	12/53
Pennsylvania Power & Light Co. Allentown, Pennsylvania Sunbury Shamokin Dam, Pennsylvania	1,000,000 125,000	1750 1525	1005 1005	Pulv. Coal And Anth.	3-D7D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	11/50	8/53
Consolidated Edison Co. Of N.Y., Inc. New York, New York East River Station New York, New York	1,400,000 180,000	2075 1800	1000 1000	Pulv. Coal	3-D7D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	3/51	6/55
Public Service Co. Of Indiana, Inc. Plainfield, Indiana Wabash River Generating Station West Terre Haute, Indiana Unit #3	700,000 100,000	1700 1500	1005 1005	Pulv. Coal	2-D7 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	5/51	8/54

TABLE V-I

CENTRAL STATION

STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulvs. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Public Service Co. of Indiana, Inc. Plainfield, Indiana Wabash River Generating Station West Terre Haute, Indiana Unit #4	700,000 100,000	1700 1500	1005 1005	Pulv. Coal	2-D7 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	5/51	12/54
Duquesne Light Co. Pittsburgh, Pennsylvania Frank R. Phillips P.S. Wireton, Pennsylvania BLR Unit #6	1,060,000 125,000	2075 1850	1000 1000	Pulv. Coal	2-D8D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	9/52	1/56
Louisville Gas & Electric Co. Louisville, Kentucky Cane Run Station Louisville, Kentucky Unit #2	775,000 103,000	1800 1590	1000 1000	Pulv. Coal & Nat. Gas	3-D6D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	1/53	1/56
Kansas Gas & Electric Co. Wichita, Kansas Murray Gill Wichita, Kansas Unit #3	800,000 100,000	2150 1925	1005 1005	Nat. Gas And Oil		Bal. Draft	Nat. Circ.		No	8/54	4/56
Kansas Gas & Electric Co. Wichita, Kansas Murray Gill Wichita, Kansas Unit #4	800,000 100,000	2150 1925	1005 1005	Nat. Gas And Oil		Bal. Draft	Nat. Circ.	Dry Ash	No	5/55	2/59

TABLE V-I

CENTRAL STATION

STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulvs. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Tennessee Valley Authority Knoxville, Tennessee Johnsonville Johnsonville, Tennessee Unit #7	1,100,000 150,000	2300 2035	1050 1003	Pulv. Coal	2-D8C EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	10/58
Tennessee Valley Authority Knoxville, Tennessee Johnsonville Johnsonville, Tennessee Unit #8	1,100,000 150,000	2300 2035	1050 1003	Pulv. Coal	2-D8C EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	1/59
Tennessee Valley Authority Knoxville, Tennessee Johnsonville Johnsonville, Tennessee Unit #9	1,100,000 150,000	2300 2035	1050 1003	Pulv. Coal	2-D8C EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	5/59
Tennessee Valley Authority Knoxville, Tennessee Johnsonville Johnsonville, Tennessee Unit #10	1,100,000 150,000	2300 2035	1050 1003	Pulv. Coal	2-D8C EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	7/59
Louisville Gas & Electric Co. Louisville, Kentucky Cane Run Station Louisville, Kentucky Unit #3	950,000 131,000	2210 1876	1005 1005	Pulv. Coal & Nat. Gas	3-D7A EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12'55	6/58

TABLE V-I

CENTRAL STATION

STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Ohio Edison Co. Akron, Ohio W. H. Sammis Plant Stratton, Ohio Unit #1	1,250,000 185,000	2700 2450	1050 1000	Pulv. Coal	5-CE No. 633 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	5/59
Ohio Edison Co. Akron, Ohio W. H. Sammis Plant Stratton, Ohio Unit #2	1,250,000 185,000	2700 2450	1050 1000	Pulv. Coal	5-CE No. 633 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	5/60
West Penn Power Co. Greensburg, Pennsylvania Armstrong Reesedale, Pennsylvania Unit #1	1,200,000 163,200	2250 1875	1010 1010	Pulv. Coal	2-D8D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/55	4/58
West Penn Power Co. Greensburg, Pennsylvania Armstrong Reesedale, Pennsylvania Unit #2	1,200,000 163,200	2250 1875	1010 1010	Pulv. Coal	2-D8D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	11/56	6/59
Ohio Edison Co. Akron, Ohio W. H. Sammis Plant Stratton, Ohio Unit #3	1,250,000 185,000	2700 2450	1050 1000	Pulv. Coal	5-CE No. 633 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	11/56	5/61

TABLE V-I

CENTRAL STATION

STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Ohio Edison Co. Akron, Ohio W. H. Sammis Plant Stratton, Ohio Unit #4	1,250,000 185,000	2700 2450	1050 1000	Pulv. Coal	5-CE No. 633 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	11/56	10/62
Empresa Nacional De Electricidad Central Termica De Compostilla II Ponferrada Spain Unit #1	945,000 147,200	2150 1900	1005 1005	Pulv. Anth.	4-D8B EX	Bal. Draft	Nat. Circ.	Dry Ash	No	2/57	5/61
Centrales Termicas Del Norte De Espana Guardo Power Plant Velilla, Spain	1,000,000 150,000	2150 1815	1005 1005	Pulv. Coal	4-D9F Press	Bal. Draft	Nat. Circ.	Dry Ash	No	6/61	6/64
Pasadena, Claifornia Pasadena Municipal Light & Power Dept. Pasadena, California Unit #1	500,000	2100 1850	1000 1000	Oil & Nat. Gas		Press	Nat. Circ.		No	12/61	6/65
Empresa Nacional De ElectricidadS . A. Centrales Termica De Compostilla II Ponferrada, Spain Unit #2	945,000 140,000	2150 1900	1005 1005	Pulv. Coal	4-D8F EX	Bal Draft	Nat. Circ.	Dry Ash	No	4/63	8/65

TABLE V-I

CENTRAL STATION

STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Tucson Gas & Electric Co. Tucson, Arizona Irvington Station Tucson Arizona Unit #4	1,140,000 158,000	2275 1975	1010 1010	Nat. Gas And Oil		Press	Nat. Circ.		Mono Wall	5/63	9/67
Nevada Power Co. Las Vegas, Nevada Reid Gardner Moapa, Nevada Unit #1	845,000 100,000	2225 1980	1005 1005	Pulv. Coal	2-D7D Press	Press	Nat. Circ.	Dry Ash	Mono Wall	5/63	5/65
Orange & Rockland Utilities, Inc. Nyack, New York Lovett Station Tomkins Cove, New York Unit #4	1,200,000 167,000	2250 1890	1000 1000	Pulv. Coal	2-D8C Press	Press	Nat Circ.	Dry Ash	Mono Wall	7/63	7/66
United Power Association Elk River, Minnesota Stanton Stanton, North Dakota Unit #1	1,200,000 172,000	2275 1875	1005 1005	Pulv. Lignite	3-MB23 Press	Bal. Draft	Nat. Circ.	Dry Ash	Mono Wall	7/64	10/66
Hidroelectria De Cataluna Barcelona, Spain Besos Power Plant Unit #1	1,060,000 160,000	2450 2170	1004 1004	Oil		Press	Nat. Circ.		Mono Wall	12/64	9/67

TABLE V-I

CENTRAL STATION

STEAM GENERATORS

Stations of Less Than 200 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Nevada Power Co. Las Vegas, Nevada Reid Gardner Station Moapa, Nevada Unit #2	845,000 100,000	2225 1980	1005 1005	Pulv. Coal	2-D7D Press	Press	Nat. Circ.	Dry Ash	Mono Wall	6/65	11/67
Hydorelectrica Moncabril SA Madrid, Spain Narcea Thermal Plant Narcea, Spain Unit #2	1,014,300 140,000	2150 1890	1005 1005	Pulv. Anth.	4-D9 Press	Press	Nat. Circ.	Dry Ash	Mono Wall	3/66	12/68

TABLE V-I

CENTRAL STATION
STEAM GENERATORS

Stations of 200 to 500 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Detroit Edison Co. Detroit, Michigan River Rouge River Rouge, Michigan Unit #3	2,000,000 321,500	2725 2450	1050 1000	Pulv. Coal	6-Riley No. 550 Duplex EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/54	10/58
Union Electric Co. St. Louis, Missouri Meramec St. Louis County, Missouri Unit #3	1,850,000 250,000	2450 2150	1010 1010	Pulv. Coal	3-D8D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	9/55	11/58
Public Service Electric & Gas Co. Newark, New Jersey Bergen Generating Station Ridgefield, New Jersey Unit #1	1,900,000 300,000	2725 2400	1100 1050	Pulv. Coal And Oil	3-D8D Press	Press	Nat. Circ.	Slag Tap	No	3/56	4/59
Public Service Electric & Gas Co. Newark, New Jersey Bergen Generating Station Ridgefield, New Jersey Unit #2	1,900,000 300,000	2725 2400	1100 1050	Pulv. Coal And Oil	3-D8D Press	Press	Nat. Circ.	Slag Tap	No	3/56	2/60
Texas Power & Light Co. Dallas, Texas Lake Creek Riesel, Texas Unit #2	1,600,000 219,000	2200 1935	1005 1005	Nat. Gas And Oil		Press	Nat. Circ.		Mono Wall	7/56	6/59

TABLE V-I

CENTRAL STATION
STEAM GENERATORS
Stations of 200 to 500 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O.T.U	Slag Tap Or Dry Ash	Mono Wall		
Wisconsin Electric Power Co. Milwaukee, Wisconsin Oak Creek Oak Creek, Wisconsin Unit #5	1,780,000 275,000	2625 2425	1050 1000	Pulv. Coal	2-D8D EX	Bal. Draft	Nat. Circ.	Dry Ash	No	12/56	12/59
Texas Electric Service Co. Fort Worth, Texas Graham Graham, Texas Unit #1	1,550,000 247,775	2550 2125	1005 1005	Nat. Gas And Oil		Press	Nat. Circ.		Mono Wall	1/57	11/60
Public Service Electric & Gas Co. Newark, New Jersey Merger Generating Station Hamilton Township, New Jersey Unit #1	2,060,000 330,000	2735 2455	1100 1050	Pulv. Coal And Oil	3-D8F Press	Press	Nat. Circ.	Slag Tap	No	6/57	12/60
Houston Lighting & Power Co. Houston, Texas W. A. Parish Richmond, Texas Unit #3	1,950,000 275,000	2525 2225	1005 1005	Nat. Gas And Oil		Press	Nat. Circ.		Mono Wall	6/57	1/61
Union Electric Co. St. Louis, Missouri Meramec Unit #4	2,310,000 300,000	2500 2200	1010 1010	Pulv. Coal	3-D9 EX	Bal. Draft	Nat. Circ.	Dry Ash	No	2/58	6/61
Public Service Electric & Gas Co. Newark, New Jersey Merger Generating Station Hamilton Township, New Jersey Unit #2	2,060,000 330,000	2734 2455	1100 1050	Pulv, Coal And Oil	3-D8F Press	Press	Nat. Circ.	Slag Tap	No	4/58	5/61

TABLE V-I

CENTRAL STATION
STEAM GENERATORS
Stations of 200 to 500 MW Capacity

UTILITY	STEAM CONDTIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Florida Power & Light Co. Miami, Florida Riviera Steam Plant Riviera, Florida Unit #3	1,950,000 300,000	2525 2100	1005 1005	Oil And Nat. Gas		Bal. Draft	Nat. Circ.		Mono Wall	8/59	5/62
Florida Power & Light Co. Miami, Florida Riviera Steam Plant Riviera, Florida Unit #4	1,950,000 300,000	2525 2100	1005 1005	Oil And Nat. Gas		Bal. Draft	Nat. Circ.		Mono Wall	2/60	3/63
Wisconsin Electric Power Co. Milwaukee, Wisconsin Oak Creek Oak Creek, Wisconsin Unit #6	1,780,000 275,000	2625 2425	1050 1000	Pulv. Coal	2-D8F EX	Bal. Draft	Nat. Circ.	Dry Ash	No	11/58	12/61
Florida Power & Light Co. Miami, Florida Port Everglades, Florida Unit #3	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat. Circ.		Mono Wall	6/61	6/64
Florida Power & Light Co. Miami, Florida Port Everglades, Florida Unit #4	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat. Circ.		Mono Wall	6/61	4/65
Florida Power & Light Co. Miami, Florida Cape Kennedy Cape Kennedy, Florida Unit #1	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat. Circ.		Mono Wall	6/61	4/65

TABLE V-I

CENTRAL STATION
STEAM GENERATORS

Stations of 200 to 500 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SER V. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O.T.U.	Slag Tap Or Dry Ash	Mono Wall		
Arizona Public Service Co. Phoenix, Arizona Four Corners Plant Fruitland, New Mexico Unit #3	1,613,000 225,000	2400 2125	1005 1005	Pulv. Coal	3-D9F Press	Bal. Draft	Nat. Circ.	Dry Ash	Mono Wall	7/61	4/64
Louisiana Power & Light New Orleans Little Gypsy Steam Electric Generating Station LaPlace, Louisiana Unit #2	2,775,000 406,500	2850 2525	1005 1005	Nat. Gas		Press	Nat. Circ.		Mono Wall	3/63	12/65
Florida Power & Light Co. Miami, Florida Turkey Point Turkey Point, Florida Unit #1	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat. Circ.		Mono Wall	8/63	11/66
Florida Power & Light Co. Miami, Florida Turkey Point Turkey Point, Florida Unit #2	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat Circ.		Mono Wall	8/63	11/67
Florida Power & Light Co. Miami, Florida Cape Kennedy Cape Kennedy, Florida Unit #2	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat. Circ.		Mono Wall	8/63	11/68

TABLE V-I

CENTRAL STATION
STEAM GENERATORS
Stations of 200 to 500 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. or Press. Or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
S. Carolina Electric Gas Co. Columbia, South Carolina Canadys, Steam Plant Canadys, South Carolina Unit #3	1,475,000 200,000	OTU 2520	1050 1000	Pulv. Coal Oil And Nat. Gas	4-MB19 Press	Press	O. T. U.	Dry Ash	Mono Wall	9/64	5/67
Omaha Public Power District Omaha, Nebraska North Omaha Station Omaha, Nebraska Unit #5	1,600,000 230,000	2875 2600	1005 1005	Pulv. Coal	3-MB21 Press	Press	Nat. Circ.	Dry Ash	Mono Wall	1/65	5/68
Florida Power & Light Co. Miami, Florida Fort Meyers Fort Meyers, Florida Unit #2	2,640,000 400,000	2850 2590	1005 1005	Oil		Press	Nat. Circ.		Mono Wall	2/66	11/69

TABLE V-I

CENTRAL STATION
STEAM GENERATORS

Stations of Greater Than 500 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Texas Electric Service Co. Fort Worth, Texas Morgan Creek Colorado City, Texas Unit #6	3,547,000 500,000	OTU 2640	1005 1005	Nat. Gas		Press	OTU		Mono Wall	3/64	5/66
New Orleans Public Service, Inc. New Orleans, Louisiana Michoud Station New Orleans, Louisana Unit #3	3,980,000 585,000	OTU 3650	1005 1005	Nat. Gas And Oil		Press	OTU		Mono Wall	11/64	4/67
Louisiana Power & Light New Orleans, Louisana Little Gypsy Steam Electric Generating Station La Place, Louisiana Unit #3	3,980,000 585,000	OTU 3650	1005 1005	Nat. Gas And Oil		Press	OTU		Mono Wall	12/64	10/68
Public Service Electric & Gas Co. Newark, New Jersey Hudson Generating Station Jersey City, New Jersey Unit #2	4,000,000 650,000	OTU 3800	1005 1030/1055	Pulv. Coal	6-MB21 Press	Press	OTU	Dry Ash	Mono Wall	1/65	10/68
Gulf States Utilities Beaumont, Texas Willow Glen Station St. Gabriel, Louisana Unit #3	3,980,000 585,000	OTU 3850	1005 1000	Nat. Gas And Oil		Press	OTU		Mono Wall	6/65	10/68

TABLE V-I

CENTRAL STATION
STEAM GENERATORS
Stations of Greater Than 500 MW Capacity

UTILITY	STEAM CONDITIONS			FUEL	FIRING EQUIPMENT		FURNACE			ORDER DATE	SERV. DATE
Name Location Station Unit Designation	Capacity P P H & KW	Press Design & Oper.	Temp. SH & RH		No. & Size Pulv. & Press. or Exhauster	Press Or Bal. Draft	Circ. Nat. Or O. T. U.	Slag Tap Or Dry Ash	Mono Wall		
Houston Lighting & Power Co. Houston, Texas Bacliff Station Bacliff, Texas Unit #3	3,980,000 585,000	O. T. U. 3650	1005 1005	Nat. Gas		Press	OTU		Mono Wall	8/65	11/68
New York State Electric & Gas Co. Binghamton, N. Y. and Pennsylvania Electric Co. Johnstown, PA. Homer City Homer City, Pa. Unit #1	4,620,000 640,000	O. T. U. 3800	1005 1005	Pulv. Coal	6-MB21.5 Press	Press	OTU	Dry Ash	Mono Wall	12/65	5/69
New York State Electric & Gas Co. Binghamton, N. Y. and Pennsylvania Electric Co. Johnstown, Pa. Homer City Homer City, Pa. Unit #2	4,620,000 640,000	O. T. U. 3800	1005 1005	Pulv. Coal	6-MB21.5 Press	Press	OTU	Dry Ash	Mono Wall	12/65	10/70
American Electric Power Co. Inc. New York, New York Plant "X" (First Unit)	5,283,000 835,000	O. T. U. 3800	1010 1025/ 1050	Pulv. Coal	6-MB23 Press	Press	OTU	Dry Ash	Mono Wall	1/66	5/69
American Electric Power Co. Inc. New York, New York Plant "X" (Second Unit)	5,283,000 835,000	O. T. U 3800	1010 1025/ 1050	Pulv. Coal	6-MB23 Press	Press	OTU	Dry Ash	Mono Wall	1/66	5/70
American Electric Power Co. Inc. New York, New York Plant "X" (Third Unit)	5,283,000 835,000	O. T. U 3800	1010 1025/ 1050	Pulv. Coal	6-MB23 Press	Press	OTU	Dry Ash	Mono Wall	1/66	5/71

VI

SULFUR VALUE RECOVERY

1. FLUE GAS CLEANING

Following absorption of heat on the walls and tube surfaces of the boilers, the combustion gas is next relieved of the heavy load of fly-ash. Mechanical separation followed by wet scrubbing is employed in CASES A, C and D. Electric precipitators operating at 900° F serve this function for CASES B and E. The cleaned gases in each system are thereafter ready to be processed for recovery of contained sulfur oxides.

2. MARKETABLE END-PRODUCTS

Of the many proposed processing methods, Ref (LL), (SS), a few only have been worked up to commercial scale application and not all of these are of interest in the present study. Most of the recovery processes that have been commercialized in Europe, USSR and Japan will yield only ammonium sulfate, a nitrogen fertilizer, that is not much favored in the United States. Even so, the fertilizer element in this substance is nitrogen; the sulfate serves as the packaging so to speak — a low value contribution. With this exception the valuable end-products of commercial operations or proposed end-products have nearly all been sulfuric acid intermediates or sulfuric acid itself in various concentrations.

In any event, it is clear that any large quantity of sulfur value recovered from coal in the United States will be consumed ultimately as sulfuric acid, but manufacture of the acid, having regard to the cost of shipping, is not necessarily best done at the site of the high sulfur combustor. Acid intermediates, as pointed out in Part III-2, are of interest also.

3. STRONG CHEMISTRY WANTED

A recovery system should employ "strong chemistry" whereby it will be capable of high efficiency in extracting sulfur oxides from the combustor flue gas, and capable of high conversion efficiency also. If these capabilities are inherent in the process chemistry and economically applied, the recovery system will provide:

3.1 High sulfur value yield.

3.2 Adaptability to a broad range of sulfur oxide concentration in the primary gas.

3.3 Minimum risk of objectionable pollution from sulfur oxide emissions.

Here it should be noted that the merits of "strong chemistry" are bought only at the cost of a high energy requirement to disengage the SO_2 from the absorbent of the recovery system in some form that can be processed ultimately to salable sulfuric acid.

4. AVAILABLE PROCESSES

These considerations together with the criterion of "availability" (See Part I-4) appear to limit the recovery process possibilities to what is listed below:

4.1 The "Contact" Process

Chemistry:

The sulfur dioxide and oxygen in the cleaned flue gas are brought into contact (hence the name) with a solid catalyst on which they react to form sulfur trioxide; this is in turn combined with water to yield sulfuric acid. If insufficient oxygen is present in the cleaned gas, more is added, usually as air. The base reactions are:

$$SO_2 + 1/2O_2 = SO_3$$

$$SO_3 + H_2O = H_2SO_4$$

4.1.1 The "Standard" Contact Process

The process has been adapted to the treatment of SO_2 containing gases from combustion of iron pyrites and elementary sulfur, from the off-gas of sulfide ore smelting, and many SO_2 sources of lesser interest such as the decomposition of ferrous sulfate and calcium sulfate, refinery acid sludge and the like. The process may be employed on gasses containing a fraction of one percent SO_2 although until recently about 4% was considered the practical lower limit. This limit has evidently been extended as noted below.

Availability

The usual range of Contact Process application is 7 to 8% SO_2 concentration up to about 13%. Several dozen installations to recover smelter off-gas are known to operate in the lower end of this range. The usual design for acid production from sulfur or pyrite employs the high end. As many as a thousand plants of the latter type have been put in operation during the nearly 50 years of commercial success of this process. It is thus incomparably the most widely practiced and highly developed process for recovery of sulfur value from SO_2-containing gas streams.

A standard Chemico Contact Process design, modified to operate on a 6% SO_2 gas, is submitted in this study as the recovery train of CASE A later described in detail. The end-product is commercial grade 93/98% sulfuric acid. Ref (R) The process is commercially offered by Chemical Construction Corporation. Similar processes are offered by other engineering organizations.

4.1.2 The Cat-Ox Process

A modification of the Contact Process has been developed in recent years by Monsanto Envirochem Systems, Inc. and associates, especially for the recovery of SO_2 from fossil fuel combustion as a means of abating air pollution. The process is capable of treating SO_2 concentrations in the range 2% down to a few tenths of 1%. Process detail for two sizes is submitted herein to serve the SO_2 recovery functions of CASES B and E. The immediate end-product of this process is 77/80% sulfuric acid. Since this grade of acid does not enjoy broad market acceptance, a standard type of acid concentrator is included whereby the output of CASES B and E is upgraded to 93% sulfuric acid.

Availability

The process has been pilot tested on a scale of about 15 MW. Ref (MM) A commercial scale (100 MW) installation of this process is to be constructed with NAPCA sponsorship at the Wood River station of Illinois Power Co. Ref (NN) The process is now commercially offered by Monsanto.

4.2 Other Sulfuric Acid Processes

A number of schemes for conversion of dilute SO_2 to sulfuric acid depend on liquid phase oxidation and represent radical departures from the well perfected technology of the "Contact Process".

Among these, one by Lurgi and one by Hitachi absorb SO_2 on activated char from which a weak sulfuric acid can be washed off with water. While these two are reported to have been tested in small pilot plants, they appear to be not available for the purpose of this study. Other proposals for direct conversion of dilute SO_2 to sulfuric acid are still in the second generation of development.

4.3 Concentration of Weak Gases to Intermediate Levels

This type of treatment is of importance since such concentrations can bridge the gap between the dilute occurrences, say less than 3% SO_2, and the established processes for treating stronger gases.

4.3.1 Reinluft Process

Chemistry:

Sulfur dioxide is absorbed on hot activated char and converted to sulfuric acid by air and steam. Unlike the proposals of Hitachi and Lurgi, the Reinluft char is then regenerated by hot inert gas that also effectively decomposes the adsorbed sulfuric acid to SO_2 and water. The SO_2 is thus concentrated to about 25% by volume from which sulfuric acid or acid intermediates can be made.

Availability:

Three commercial installations are or have been in operation in Germany of which the largest is at Wolfsburg designed to treat 50, 000 SCFM of dilute SO_2 gas. The Reinluft process, while evidently in an advanced stage of development, is believed to be still subject to some degree of explosion hazard. The process is objectionable also in consequence of ground-area requirements that are disproportionately large, mainly because of the low space velocity of the primary adsorption. Ref (R) (EE) (LL)

4.3.2 Grillo Process

Chemistry:

Sulfur oxides in flue gas are absorbed in "mixed metal oxides", typically manganese dioxide and magnesium hydroxide on a coke carrier. The absorbent is screened out of the coke, mixed with coal and regenerated at 1,650° F to produce a sulfur-bearing gas stream that is burned to yield sulfur dioxide suitable for manufacture of acid or acid intermediates. The solid absorbent is water-quenched and recycled.

Availability:

A pilot unit of 8 MW scale was operated in Cologne, Germany by the sponsor, A.G. fur Zinkindustrie vorm Wilhelm Grillo. This process like the process of Reinluft is handicapped by large area requirements and has probably not been developed to available status. Ref. (LL) (R) (EE)

4.3.3 Magnesium Sulfite/Oxide Process

Chemistry:

Flue gas containing sulfur oxides and free of particulate matter is scrubbed with a slurry of magnesium oxide and water with which the sulfur oxides react to form a complex of hydrated magnesium sulfites, bisulfites, and sulfates when oxygen is present. The precipitate from this is dehydrated and calcined by heating with fuel gas in the presence of coke to yield a 15% SO_2 gas concentrate, together with regenerated magnesium oxide for return to the absorber. Ref. (PP) The "strong chemistry" of this process gives it capability over a wide range of SO_2 concentrations in the flue gases. Dry magnesium sulfur salts (principally $MgSO_3$) are intermediate to the 15% SO_2 gas product and can be shipped for sulfuric acid manufacture at a distance, for example to a central processing plant. The chemistry is capable of high yield of sulfur oxides treated since sulfate formation is reduced by the coke in the calcination step.

Availability:

Some aspects of this process have been in operation for a number of years in nineteen plants as an adjunct to magnesium-base sulfite pulping of wood. The chemistry exploited in the pulp industry is the somewhat weaker magnesium sulfite/bisulfite cycle but the SO_2 absorption and the magnesium sulfite thermal decomposition are analogous to features of the magnesium oxide/sulfite cycle. Ref (QQ) This auxiliary cycle for wood pulping was primarily a joint development of Babcock and Wilcox Co., the Weyerhaeuser Co., and the Canadian firm, Howard Smith Paper Mills, Ltd. Babcock and Wilcox are believed now to offer commercially for pollution control a magnesium oxide wet scrubbing system with SO_2 recovery. Ref (Y) (HH)

Control of SO_2 emissions from other sources using the magnesium oxide/sulfite slurry in Venturi absorbers has been pilot plant tested by Chemical Construction Corporation on the sulfuric acid plant emission of Olin Mathieson Company, Baltimore, Maryland, and on flue gas of Canal Electric Co., Sandwich, Massachusetts. The scale of these tests was about 2,000 SCFM. During the last several years an active design-development program has been sponsored by Chemical Construction Corporation and Basic Chemicals Division of Basic Incorporated, Cleveland, Ohio. The process is commercially offered by Chemico-Basic, a joint venture of these two firms. The Chemico-Basic process will be installed at the 150 MW Mystic station of Boston Edison Co. Ref (RR)

This process is the basis for the gas cleaning and SO_2 concentrating phase of the recovery systems of CASE C and CASE D, summarized in Part VII of this volume and presented in greater detail in Volume Two.

4.4 Concentration of Weak Gases to High (90/100%) Level

Processes of this type have usually been employed for liquid sulfur dioxide end-product. This compound can serve as intermediate to sulfuric acid manufacture but is not so far much used in this way since advantage would accrue only to the shipping weight relation: 2 tons of liquid SO_2 will make about 3 tons of acid. Liquid SO_2 enjoys a small market other than for acid making, but the outlook for a large volume of use appears to depend on increasing demand by acid manufacturers.

emission of Olin Mathieson Company, Baltimore, Maryland, and on flue gas of Canal Electric Co., Sandwich, Massachusetts. The scale of these tests was about 2,000 SCFM. During the last several years an active design-development program has been sponsored by Chemical Construction Corporation and Basic Chemicals Division of Basic Incorporated, Cleveland, Ohio. The process is commercially offered by Chemico-Basic, a joint venture of these two firms. The Chemico-Basic process will be installed at the 150 MW Mystic station of Boston Edison Co. Ref (RR)

This process is the basis for the gas cleaning and SO_2 concentrating phase of the recovery systems of CASE C and CASE D, summarized in Part VII of this volume and presented in greater detail in Volume Two.

4.4 Concentration of Weak Gases to High (90/100%) Level

Processes of this type have usually been employed for liquid sulfur dioxide end-product. This compound can serve as intermediate to sulfuric acid manufacture but is not so far much used in this way since advantage would accrue only to the shipping weight relation: 2 tons of liquid SO_2 will make about 3 tons of acid. Liquid SO_2 enjoys a small market other than for acid making, but the outlook for a large volume of use appears to depend on increasing demand by acid manufacturers.

4.4.1 Aromatic Amine Absorbents

Chemistry:

Xylidine-water mixtures and dimethyl aniline have been used as absorbents to concentrate SO_2 from dilutions of 3.5% and stronger. Absorption and desorption are accomplished on a thermal cycle. The gas concentrate is of high purity SO_2 suitable for liquefaction. The aromatic vapors are recovered by absorption in sulfuric acid and the resultant amine sulfates are treated with soda ash to release the amine. Sodium sulfate is discarded.

Availability:

American Smelting and Refining Co. installed the DMA version of this process at three smelter operations 20-25 years ago. Of these, two are believed to be still in operation (on 4% SO_2 gas) but the plants have remained small since the product is sold in the liquid SO_2 market, a minor factor in this picture. Dimethyl aniline, xylidene and several other organic materials, (dimethyl formamide, for example) that have a substantial solvent capacity for sulfur oxides may be in general less suitable than alkali/water systems for treating large volumes of hot gas at high velocities. The cost of solvent losses and possible explosion hazard are unfavorable factors. Ref (VV)

4.4.2 Basic Aluminum Sulfate

Chemistry:

This SO_2 absorption medium is made by addition of lime to aluminum sulfate solution to precipitate a portion of the SO_3 radical whereby the cool, clarified liquor gains relatively great solvent power for SO_2 which may be expelled by heating. The product approximates 100% SO_2 gas.

Availability:

This process, developed by Imperial Chemical Industries, Ltd., England, was first commercially used on the copper smelter emission of Outokumpu Oy at Imatra, Finland in 1936. Ref (SS) The product was 52 tons per day of 100% SO_2 from 5% SO_2 off gas. ICI also conducted pilot testing at Billingham, but this and the Imatra plant have long since been dismantled. The process is currently operated at Manchester, England, by Hardman & Holden, Ltd., who make 30 tons per day of liquid SO_2 from combustion of sulfur. The Manchester facility was carefully examined for applicability of this process to the CASE C and D combustor gases of this study in order to gain advantage from the highly concentrated SO_2 gas as a feed material for sulfur production. However, the basic aluminum sulphate concentrating cycle as now operated is subject to a bothersome lime-scaling of equipment, and to a considerably diminished efficiency when used on very weak gases. While this process may be the one of choice for manufacture of liquid SO_2 from an intermediate concentration, the process combinations selected for CASES C and D appear to be a better route to sulfur making. The basic aluminum sulfate absorbent may not be suitable for use directly on coal combustion gases since there is some evidence that metals such as arsenic, tellurium and antimony are harmful to it. Ref (HH) (SS)

4.4.3 Thermal Cycle With Water

Chemistry:

Cooled dilute sulfur dioxide gas is dissolved in cold water at atmospheric pressure, followed by release of the SO_2 concentrate from the solution by steaming or other heating. The dried gas is virtually 100%, and suitable for liquefaction or reduction to sulfur.

Availability:

This process is operated at about 30 tons per day SO_2 output rate by the Associated Chemical Companies Division of Albright & Wilson, Ltd. at Bromborough, England. The SO_2 is made by combustion of sulfur in air; the concentrate is liquefied. Preliminary economic study indicates that this process may be competitive with basic aluminum sulfate absorption unless the tendency of the latter to develop scale in the absorption equipment can be reduced.

4.4.4 Alkali-Metal Sulfite/Bisulfite

Chemistry:

The principal reactions of the absorption are:

$$2MOH + SO_2 \longrightarrow M_2SO_3 + H_2O$$

$$M_2SO_3 + SO_2 + H_2O \longrightarrow 2MHSO_3$$

The bisulfite can be steam-stripped to yield SO_2 of high purity. Ref (TT)

Availability:

Wellman-Lord, Inc., Lakeland, Florida are the principal contemporary proponents of this chemistry. The company was reported to favor the potassium system and to crystallize potassium pyrosulfite as an intermediate step. The pyrosulfite reverts to the bisulfite form when dissolved in water for steam stripping. The chemistry of these systems for SO_2 absorption has been studied for nearly 40 years by several investigators. Ref (LL) Wellman-Lord reports that a 3-year development program included 1 MW pilot plant at the Gannon Station of Tampa Electric Co. and a 25 MW demonstration plant at Baltimore Gas and Electric Co., Trane Station, together with test work at two other utility installations. A prototype plant of the Wellman-Lord process was started up in July, 1970 to recover sulfur oxide emissions from destruction of refinery acid sludge at Olin Mathieson's

Paulsboro, N.J., facility. The process as installed at Paulsboro is reported to use a regenerative sodium sulfite system that will recover as SO_2 concentrate not less than 90% of sulfur oxides treated, and limit the final emission to less than 500 parts per million. Operating data however have been insufficient to evaluate this process for the purpose of this study. Ref (UU)

4.5 Liquefaction of SO_2

So far as known liquid SO_2 is produced from strong (usually 90 to 100%) gas by cooling, drying, compressing and recooling to condense the liquid product. Engineering of this process is readily available from experienced sources.

An effort by Canadian Industries, Ltd. to liquefy SO_2 from dilute gas was carried into the pilot plant stage but abandoned because of high power cost. Ref (R)

4.6 Processes for Output of Elementary Sulfur

Reduction of sulfur dioxide to sulfur has been practiced for many years with commercialized technology under the incentive of pollution control leavened by income from sale of sulfur. A recent, comprehensive review of the state of the art has been reported to NAPCA by Allied Chemical Corporation, Ref (GG). The successful processes have utilized one or another (or more than one) of the four major reducing agents:

Hydrogen sulfide,

Carbon as coke,

Reformer or producer gas (mixtures of CO and H_2 derived from natural gas or coke),

Gaseous hydrocarbons, usually natural gas.

4.6.1 Reduction of 100% SO_2 by Hot Coke

Chemistry:

Sulfur dioxide from the concentrating plant, diluted with air to 50/60% SO_2 strength, and preheated to 300° C, reacts with coke qualitatively as follows:

$$C + O_2 \longrightarrow CO_2$$
$$C + SO_2 \longrightarrow CO_2 + S$$
$$C + CO_2 \longrightarrow 2CO$$
$$2CO + 2S \longrightarrow 2COS$$

Any moisture or hydrogen present will give rise to H_2S. The reactions are highly exothermic. The gases leave the coke bed at 800/900° C with some unconverted SO_2 which is thereupon reacted over a catalyst with CO, COS and H_2S to yield more sulfur. The liquid sulfur from this process is of high purity.

Availability:

The process was tested by Imperial Chemical Industries Ltd., at Billingham, England in a plant having 5 to 6 tons per day output of sulfur. Development work was completed and the unit was shut down for lack of a profitable outlook as of the late 1930's.

4.6.2 Reduction of Dilute SO_2 with Producer Gas

Chemistry:

The reducing agents in "producer gas", which is generally made from coke, are carbon monoxide and hydrogen generated according to:

$$H_2O + C \longrightarrow CO + H_2$$

Reactions of these agents with SO_2 are somewhat complex; the following are typical:

$$\begin{aligned}
SO_2 &+ 2CO &&\longrightarrow 2CO_2 &&+ S\\
SO_2 &+ 2H_2 &&\longrightarrow 2H_2O &&+ S\\
CO &+ S &&\longrightarrow COS &&\\
C &+ 2S &&\longrightarrow CS_2 &&\\
H_2 &+ S &&\longrightarrow H_2S &&\\
SO_2 &+ 2H_2S &&\longrightarrow 2H_2O &&+ 3S\\
SO_2 &+ 2COS &&\longrightarrow 2CO_2 &&+ 3S\\
SO_2 &+ CS_2 &&\longrightarrow CO_2 &&+ 3S
\end{aligned}$$

A principal process exploiting this kind of chemistry generates producer gas by diluting with wet air a portion of the raw SO_2-containing smelter gas that is to be treated, passing this mixture through red hot coke, recombining the remainder of the raw gas with the coke-reacted gas, and flowing the whole over a catalyst to promote the equilibrium products which are mainly carbon dioxide, water and sulfur, with small amounts of unreduced sulfides that were vented in early practice. This tail gas can be oxidized and recycled in favorable circumstances. See Part VI 5. 1.

Availability:

The process was developed by Bolidens Gruvaktiebolag, and operated in their smelter works at Rönnskär, Sweden, to make 25, 000 tons of sulfur annually beginning in 1935. Some years later, it was replaced by a process producing liquid SO_2 from flue gases of iron pyrite roasting. Ref (SS)

A similar process was operated for output of 150 tons per day of sulfur from 1935 to 1943 by Consolidated Mining and Smelting Co. of Canada, Ltd., at Trail, B.C. The plant was shut down for market reasons in favor of direct oxidation of SO_2 to ammonium sulfate and sulfuric acid. Ref (WW)

4. 6. 3 Reduction of SO_2 with Methane

Chemistry:

4. 6. 3. 1 A non-catalytic process (ASARCO) has been developed for the treatment of hot smelter off-gas containing 5 to 7% SO_2 and 12 to 9% O_2. Methane sufficient for reduction of the SO_2 and the O_2 is added to the off-gas as it enters a brick checker-

work filled combustion chamber in which the principal reaction yields carbon dioxide, water and sulfur:

$$CH_4 + O_2 + SO_2 \longrightarrow CO_2 + 2H_2O + S$$

Hydrogen sulfide and carbonyl sulfide occur from side-reactions and are reduced to sulfur in downstream catalytic reactors. Sulfur recovery is reported to be 95% in total. Ref (GG) (R) (XX)

4.6.3.2 Catalytic processes, especially that of West (TGS), employ a direct reaction of methane and SO_2 over a catalyst. A relatively low temperature is favorable, and oxygen accompanying the SO_2 is proportionately objectionable. Note that the 15% SO_2 concentrate of CASE C and D contains very little oxygen and is thus adapted to the type of catalyzed primary reaction proposed by West and Yushkevitch (low space velocities) and Kulcsar (high space velocity). Ref (GG) The process employed in the systems of CASE C and D differs from these in removing the reform reaction to a separate zone, and in some other particulars. (See Part VI 4.6.4)

Availability:

The non-catalytic process (4. 6. 3. 1) was operated from 1940 to 1944 on a semicommercial scale by American Smelting and Refining Co. High quality sulfur was produced. Ref (XX) Interest declined for lack of a profitable sulfur market.

The catalytic process of West was developed by Texas Gulf Sulfur Co. and International Nickel Company in a pilot plant. Ref (R)

4. 6. 4 Reduction of Dilute SO_2 with Reformed Natural Gas (Methane)

Chemistry:

Since the reducing agents in gas from "reformed" hydrocarbons are carbon monoxide and hydrogen, as in producer gas, the reactions with SO_2 are qualitatively similar. (See Part VI, 4. 6. 2) The equilibrium products are carbon dioxide, water and sulfur with minor amounts of volatile sulfides. In a manner similar to the producer gas process, these gases are reacted on catalysts to accelerate arrival at the equilibrium conditions. The catalyzed reaction of SO_2 with reformed natural gas is the basis of sulfur recovery from 15% gas concentrate in CASES C and D and is described in more detail in Volume Two. In this adaptation the initial reduction yields a mixture of sulfides, principally H_2S, which together with SO_2 are thereupon treated in a Claus reactor. Sulfides remaining in the tail gas are reoxidized to SO_2 and returned to the magnesium sulfite/oxide cycle.

Availability:

This process and catalyst have been developed by Chemical Construction Corporation and pilot tested at a scale of 25 pounds of sulfur per hour together with life-testing of the catalyst. Similar work has been carried out by Allied Chemical Corporation as part of a comprehensive review of all known SO_2 reduction chemistry. Ref (GG)

The reduction of SO_2 with reformed natural gas is reported by Allied to show "good, and very similar economics" when compared to the ASARCO high temperature methane reduction. Allied have installed a process for sulfur recovery from smelter gases at Sudbury, Onatrio, Canada, which is believed to employ reformed natural gas as the reductant for 180, 000 short tons per year of sulfur output.

4. 6. 5 Reduction of SO_2 with H_2S

Chemistry:

The catalyzed reaction $2H_2S + SO_2 \rightarrow 2H_2O + 3S$ is frequently employed on sources of H_2S. Usually a portion of the H_2S concentrate is burned in air to SO_2 which is remixed with H_2S in the required ratio. This process is named after the inventor, Claus. When the primary source is SO_2, a portion of it must be reduced to H_2S to prepare the Claus feedstock, and if carbon or hydrocarbon is used in this reduction the occurrence of carbon disulfide (CS_2) and carbonyl sulfide (COS) is to be expected. In a manner analogous to Claus these compounds are catalytically reacted with SO_2 to form sulfur. Several variants of Claus techniques have been developed to serve particular upstream conditions. All processes for reduction of SO_2 to sulfur appear to require some application of Claus chemistry.

Availability:

More than 200 Claus units have been put in operation, for the most part in desulfurization of petroleum fractions and "sour" natural gas. Engineering services and catalyst for Claus units are offered by several firms. Claus units are included in the systems of CASE C and CASE D in this study.

5. EMISSION CONTROL

5.1 All sulfur oxide emissions from the five high sulfur combustor systems analyzed for purposes of this study are limited by design to less than 100 ppm of volume of total tail gas emission. CASE C and CASE D systems provide for recycle of gaseous sulfur compounds from the Claus process exit, through the "incinerator" oxidation, back to the flue gas scrubber and hence to the main absorber. However, a very low sulfur fuel, for example, pipeline gas is needed for the magnesium salts drying operation since the emission of the dryers goes directly to the main stack for reheat purposes. See Part XIV.

The systems of CASE A, B and E include small venturi absorbers, especially provided for tail gas cleanup, but charged with calcium oxide instead of magnesium oxide.

5.2 Stack heights and stack gas temperatures included in the designs are intended to limit maximum 24-hour downwind ambient sulfur oxide concentrations to 0.1 PPM.

5.3 All particulate emissions are subject in the designs to recommended criteria for control of interstate air pollution in the Parkersburg, West Virginia-Marietta, Ohio area as issued by Commissioner John T. Middleton of the National Air Pollution Control Administration on March 23, 1970.

VII
SUMMARY OF HSC SYSTEMS

1. SUMMARY OF THE CASE A SYSTEM

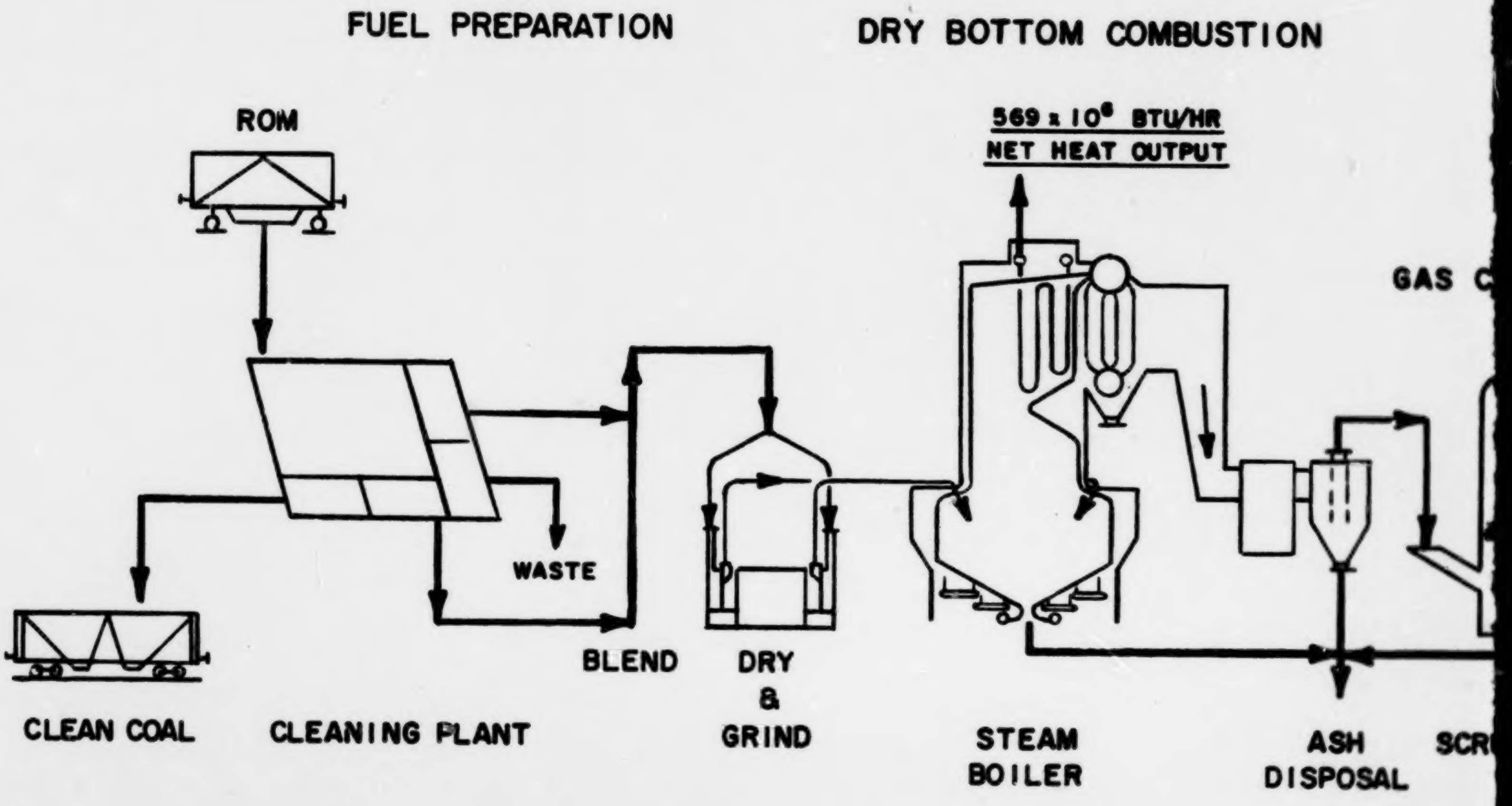
(Refer to diagram following)

A run of mine bituminous coal containing 3 to 4.5% sulfur, washable to 1% sulfur, is assumed. The reject is treated for maximum yield of an HSC Fuel of the CASE A specification. This fuel is pulverized and fired to raise 500,000 lbs/hr of industrial quality steam and a flue gas containing 6% sulfur dioxide and 4.3% oxygen. The gas is wet cleaned in a packed tower, dried and converted catalytically to sulfur trioxide which is reacted with water to make 98% sulfuric acid at the rate of 1,830 net tons (100% basis) per day, 330 days per year. Sulfur emission equals 0.4% of sulfur charged. Sulfur not emitted equals 93 pounds per million Btu of heat recovered in the CASE A System, or 2.2 pounds per million Btu of ROM coal charged to the cleaning plant.

CASE A is described in greater detail in Volume Two, Part XIII.

a

HIGH SULFUR COM


FUEL PREPARATION
DRY BOTTOM COMBUSTION
GAS C
ROM
569 x 10⁶ BTU/HR
NET HEAT OUTPUT
WASTE
CLEAN COAL
CLEANING PLANT
BLEND
DRY & GRIND
STEAM BOILER
ASH DISPOSAL
SCR

FUELS COMPOSITION
WT. % DRY BASIS*

	RUN OF MINE	CLEAN COAL	HSC FUEL
IRON PYRITE			61.6
ORGANIC SULFUR		1.0	0.2
TOTAL SULFUR	3 TO 4.5	1.0	(33.1)
COAL COMBUSTIBLES			18.2
NON-PYRITIC ASH			20.0
M TONS/YEAR			640.0
TONS/HR: IRON PYRITE			49.3
COAL COMBUSTIBLES			14.7
NON-PYRITIC ASH			16.0
SULFUR/COAL RATIO			1.8
BTU/LB			4420
BTU/HR			700×10^6

*6% MOISTURE ASSUMED IN HSC FUEL AS FIRED

GAS C
VOL. %

	EXIT FURNA
SO_2	
SO_3	
O_2**	
CO_2	
N_2	
H_2O	4.0
TEMP. °F	500
M SCFM	177
ASH:	
TONS/HR	48.8
% FE	47

** 25% EXCESS COMBUST

B

LFUR COMBUSTOR STUDY

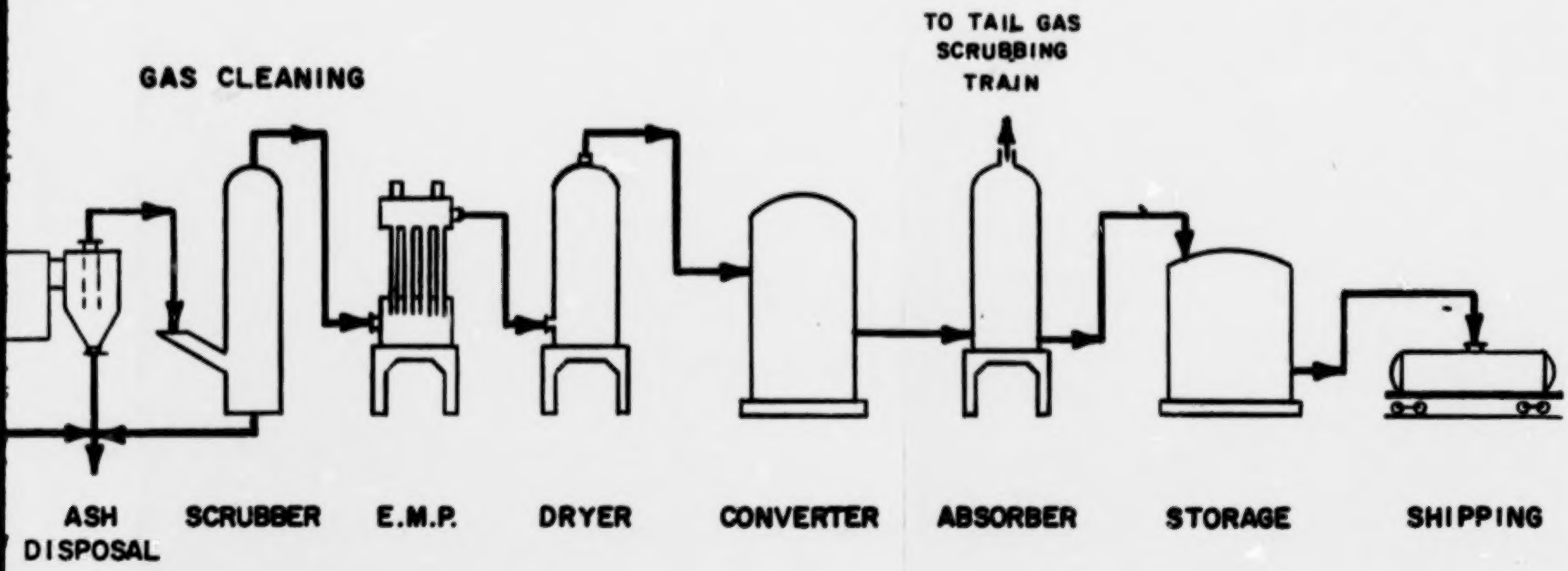
CASE A

ENERGY PRODUCT STEAM

500,000 LBS/HR
825° F 900/PSIG

ION

SULFURIC ACID PLANT
(2 TRAINS)


GAS CLEANING
TO TAIL GAS SCRUBBING TRAIN
ASH DISPOSAL
SCRUBBER
E.M.P.
DRYER
CONVERTER
ABSORBER
STORAGE
SHIPPING

GAS COMPOSITION VOL. % DRY BASIS

EXIT FURNACE	FEED TO CONVERTERS
6.0	5.0
0.2	
4.3	7.1
8.6	7.2
80.9	80.7
4.0	
500	
177	203
48.8	
47	

EXCESS COMBUSTION AIR

CHEMICAL PRODUCT

COMMERCIAL GRADE 98 % SULFURIC ACID

1830 TONS/DAY
100 % BASIS

2. SUMMARY OF THE CASE B SYSTEM

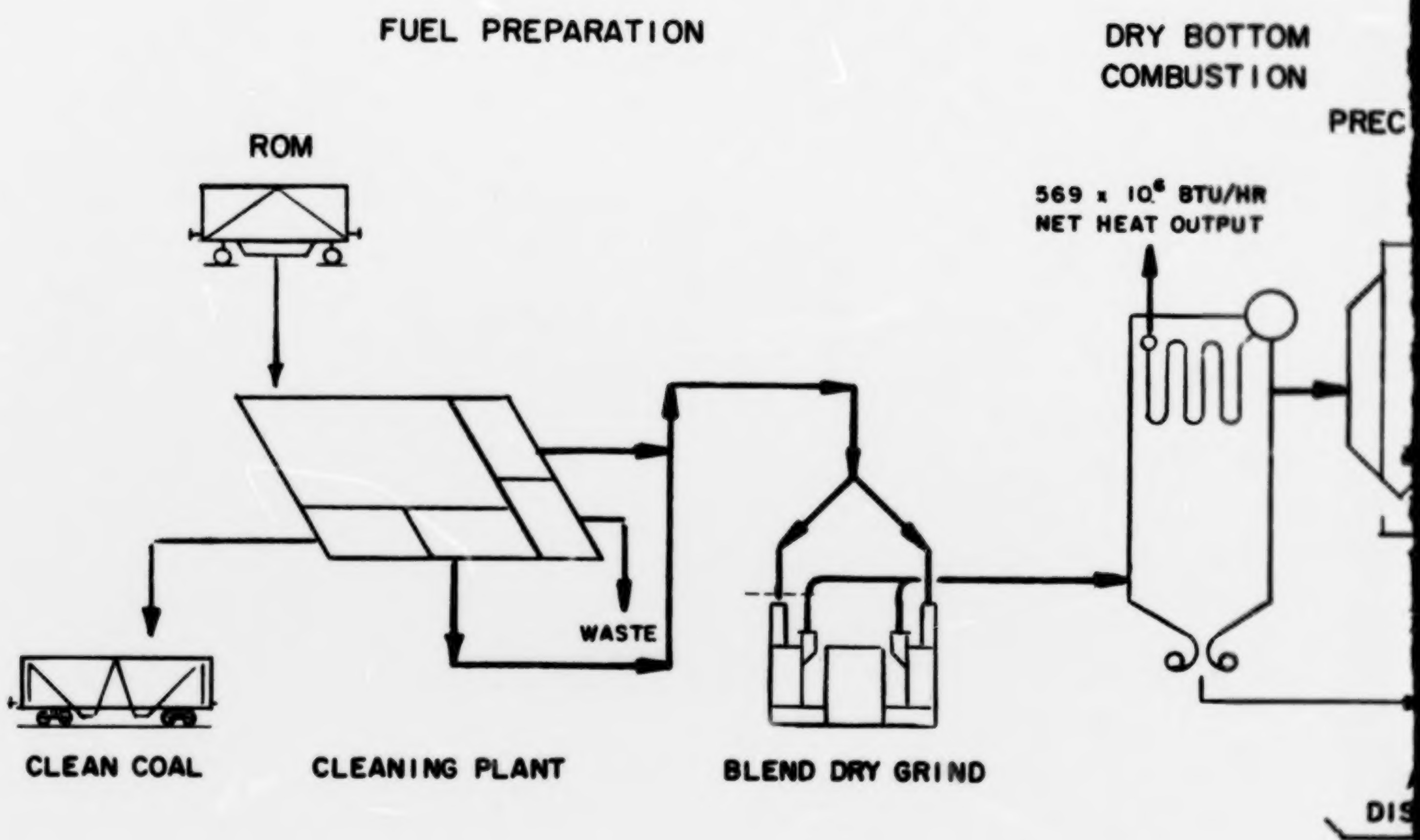
(Refer to diagram following)

A run of mine bituminous coal containing 3 to 4.5% sulfur, washable to 1% sulfur, is assumed. The reject is treated for maximum yield of an HSC Fuel of the CASE B Specification. This fuel is pulverized and fired to raise 500,000 lbs/hr of industrial quality steam and a flue gas containing 2% sulfur dioxide and 4.2% oxygen. The gas is cleaned of particulates in an electric precipitator at 805° F and is then converted catalytically to sulfur trioxide which is reacted with water to make 77/80% sulfuric acid by the Cat-Ox process of Monsanto at the rate of 624 net tons (100% basis) per day, 330 days per year. Sulfur emission equal 0.5% of sulfur charged. Sulfur not emitted equals 33 pounds per million Btu of heat recovered in the CASE B System, or 2.0 pounds per million Btu of ROM coal charged to the cleaning plant.

CASE B is described in greater detail in Volume Two, Part XIV.

a

HIGH SULFUR COMBUSTO


FUEL PREPARATION
DRY BOTTOM
COMBUSTION
PREC
ROM
569 x 10^6 BTU/HR
NET HEAT OUTPUT
WASTE
CLEAN COAL
CLEANING PLANT
BLEND DRY GRIND
DIS

FUELS COMPOSITION
WT % DRY BASIS*

	RUN OF MINE	CLEAN COAL	HSC FUEL
IRON PYRITE			33.8
ORGANIC SULFUR			0.5
TOTAL SULFUR	3 TO 4.5		(18.5)
COAL COMBUSTIBLES			45.7
NON PYRITIC ASH			20.0
M TONS/YEAR			406
TONS/HR: FeS$_2$			17.1
COAL			23.4
N-P ASH			10.1
SULFUR/COAL RATIO			0.4
BTU/LB			7400
BTU/HR			750 x 10^6

* 6% MOISTURE ASSUMED IN HSC FUEL AS FIRED

GAS COM
VOL % D

SO$_2$
SO$_3$
O$_2$
CO$_2$
N$_2$
H$_2$O
TEMP
M ACF
M SCF
ASH
TON
% F

COMBUSTOR STUDY

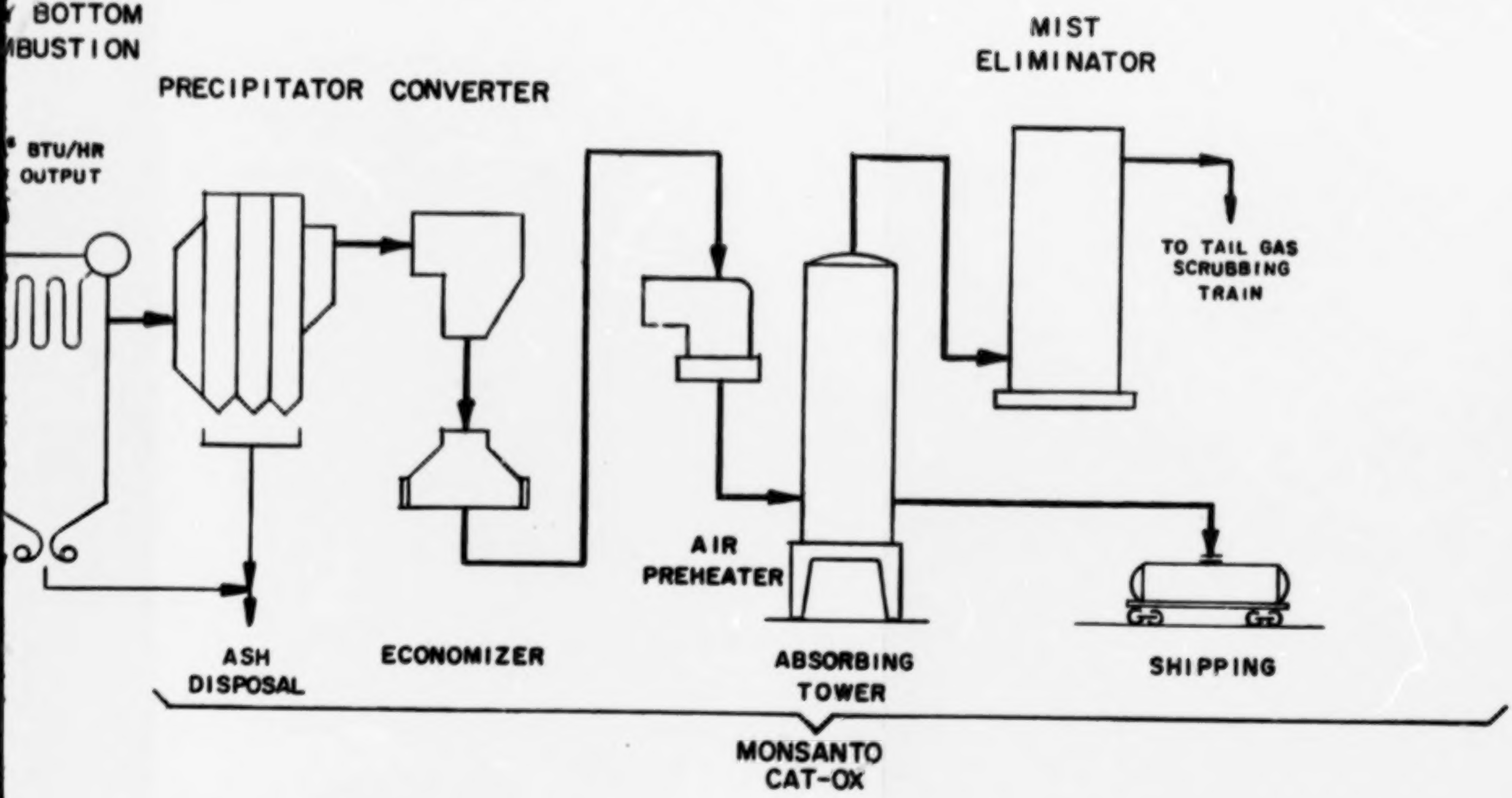
B

CASE **B**

ENERGY PRODUCT

STEAM

500,000 LBS/HR
825 F 900 PSIG


BOTTOM
BUSTION
BTU/HR
OUTPUT
PRECIPITATOR
CONVERTER
MIST ELIMINATOR
TO TAIL GAS SCRUBBING TRAIN
AIR PREHEATER
ASH DISPOSAL
ECONOMIZER
ABSORBING TOWER
SHIPPING
MONSANTO CAT-OX

GAS COMPOSITION VOL % DRY BASIS

	EXIT FURNACE
SO_2	2.0
SO_3	0.1
O_2	4.2
CO_2	14.0
N_2	79.7
H_2O	**
TEMP °F	805
M ACFM	445
M SCFM	173
ASH	
TON/HR	21.6
% FE	37

** NOMINAL MOISTURE CONTENT 4%

CHEMICAL PRODUCT

COMMERCIAL
GRADE 77/80%
SULFURIC ACID

624 TONS/DAY
100% BASIS

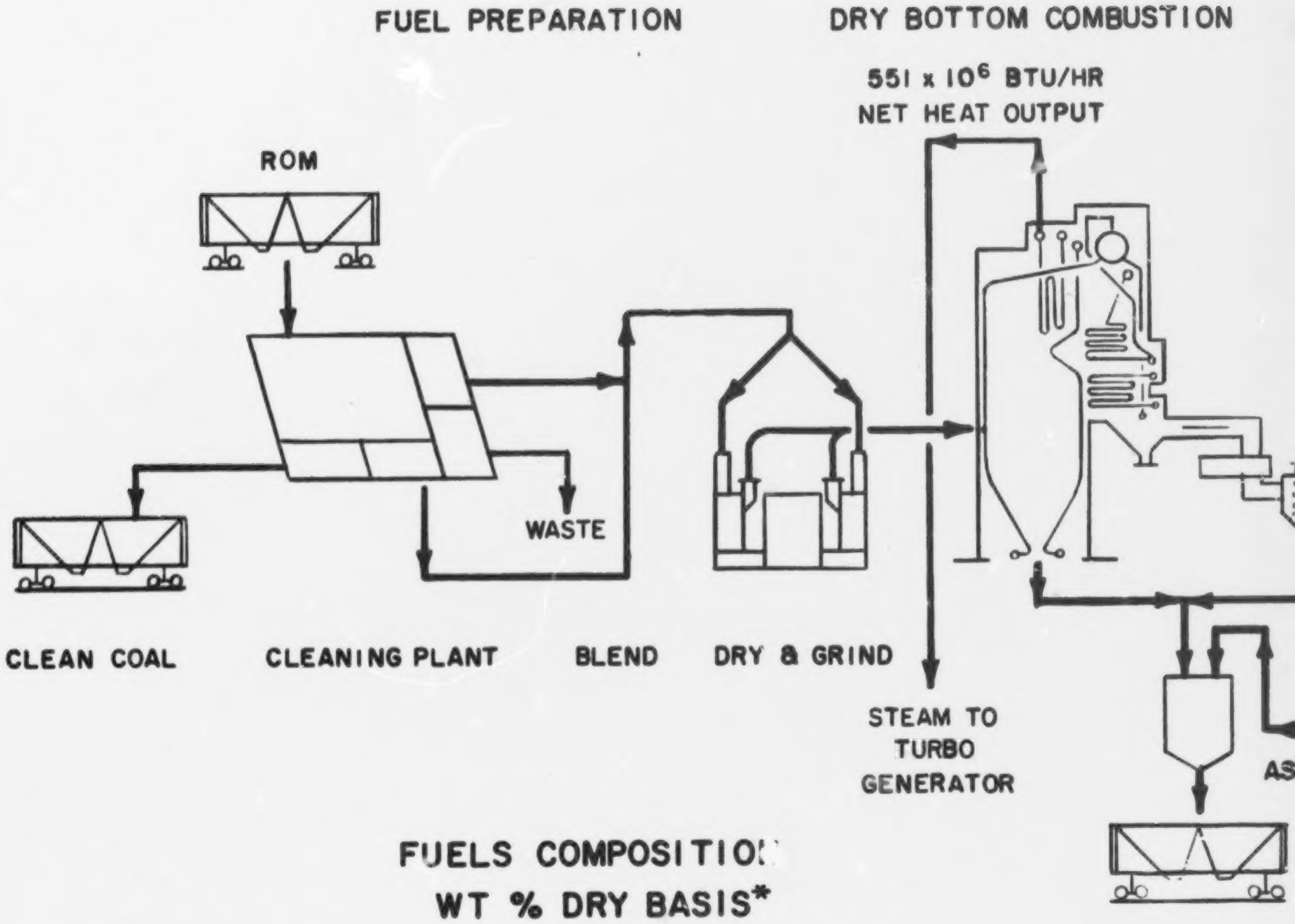
3. SUMMARY OF THE CASE C PROTOTYPE SYSTEM

(Refer to diagram following)

A run of mine bituminous coal containing 3 to 4.5% sulfur, washable to 1% sulfur, is assumed. The reject is treated for maximum yield of an HSC Fuel of the CASE C and D specification. This fuel is pulverized and fired to raise 500,000 lbs/hr of subcritical steam which is fed to a turbogenerator of 50 MW capacity. The flue gas containing about 1% sulfur dioxide and 4.3% oxygen is wet-cleaned in a venturi scrubber and led to a two-stage-venturi absorber in which sulfur dioxide is reacted, essentially with magnesium oxide, to produce magnesium sulfite. This salt is separated, dried, and calcined to yield a 15% sulfur dioxide gas concentrate and magnesium oxide which is recycled. The 15% sulfur dioxide is converted by catalytic reduction to sulfur at the rate of about 100 net tons per day, 330 days per year. Sulfur emission equals 1.3% of sulfur charged. Sulfur not emitted equals 16 pounds per million Btu of heat recovered in the CASE C Prototype System, or 2.2 pounds per million Btu of ROM coal charged to the cleaning plant.

CASE C is described in greater detail in Volume Two, Part XV.

a


FUEL PREPARATION
DRY BOTTOM COMBUSTION
551 x 10⁶ BTU/HR
NET HEAT OUTPUT
ROM
WASTE
CLEAN COAL
CLEANING PLANT
BLEND
DRY & GRIND
STEAM TO
TURBO
GENERATOR
AS

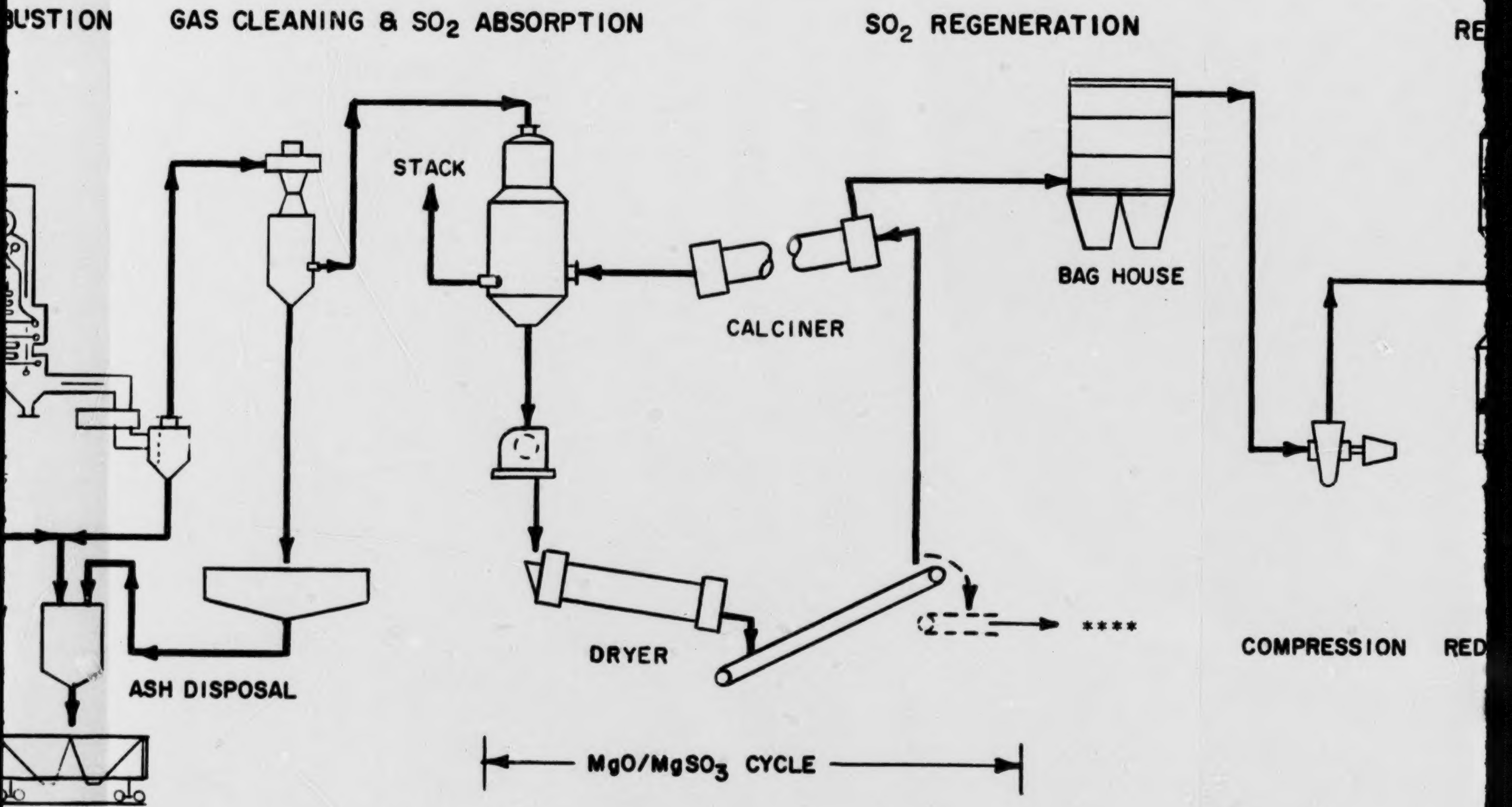
FUELS COMPOSITION
WT % DRY BASIS*

	RUN OF MINE	CLEAN COAL	HSC FUEL
IRON PYRITE			20.8
ORGANIC SULFUR		1.0	0.6
TOTAL SULFUR	3 TO 4.5		(11.7)
COAL COMBUSTIBLES			58.6
NON-PYRITIC ASH			20.0
MM TONS/YEAR			298.0
TONS/HR: IRON PYRITE			7.7
COAL COMBUSTIBLES			22.0
NON PYRITIC ASH			7.4
SULFUR/COAL RATIO			0.2
BTU/LB			9000
BTU/HR			670×10^6

* 6% MOISTURE ASSUMED IN HSC FUEL AS FIRED

B

HIGH SULFUR COMBUSTOR STUDY


USTION
GAS CLEANING & SO_2 ABSORPTION
SO_2 REGENERATION
RE
STACK
BAG HOUSE
CALCINER
DRYER
ASH DISPOSAL
****
COMPRESSION
RED
MgO/$MgSO_3$ CYCLE

HSC UEL
20.8
0.6
(1.7)
58.6
20.0
98.0
7.7
22.0
7.4
0.2
00
0 x 10^6

FLUE GAS VOL % DRY BASIS**		ASH COMPOSITION		CALCINER EMISSION	
SO_2	1.15	%FE	29	SO_2	15.9
SO_3	0.05	TONS/HR	12.6		
O_2	4.3***				
CO_2	15.4			CO_2	13.6
N_2	79.1			N_2	70.5

TEMP °F	500
MACFM	308
MSCFM	167

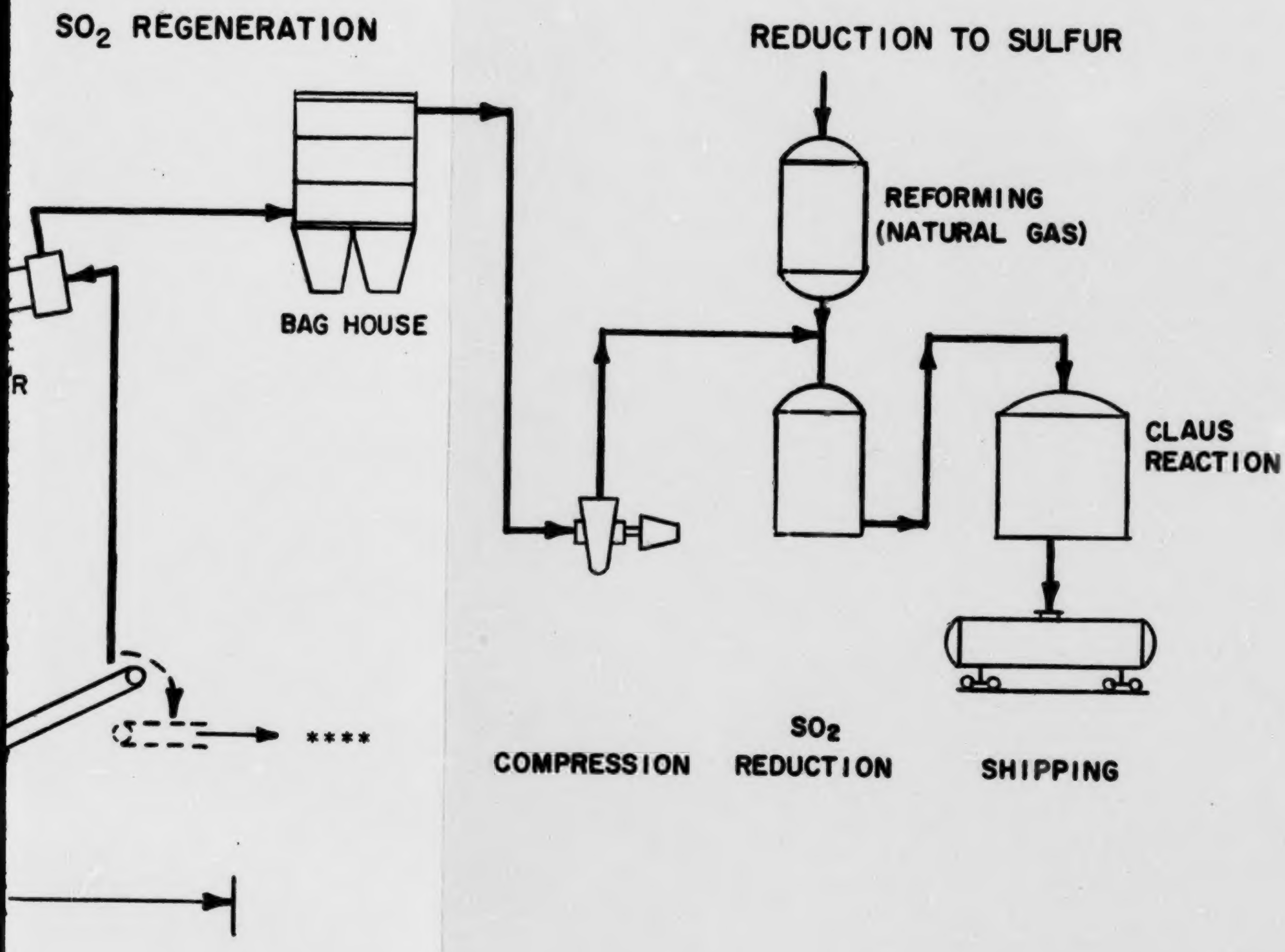
**NOMINAL MOISTURE CONTENT 4%

***25% EXCESS COMBUSTION AIR

****MAGNESIUM SALTS MAY BE SHIPPED BEFORE PROCESSING

C

TUDY


SO_2 REGENERATION
REDUCTION TO SULFUR
REFORMING (NATURAL GAS)
BAG HOUSE
R
CLAUS REACTION
****
COMPRESSION
SO_2 REDUCTION
SHIPPING

CALCINER EMISSION

SO_2	15.9
CO_2	13.6
N_2	70.8

****MAGNESIUM SALTS MAY BE SHIPPED BEFORE PROCESSING

PROTOTYPE SYSTEM

CASE C

ENERGY PRODUCT

ELECTRICITY
50MW CAPACITY

CHEMICAL PRODUCT

COMMERCIAL GRADE

SULFUR

102 NET TONS/DAY

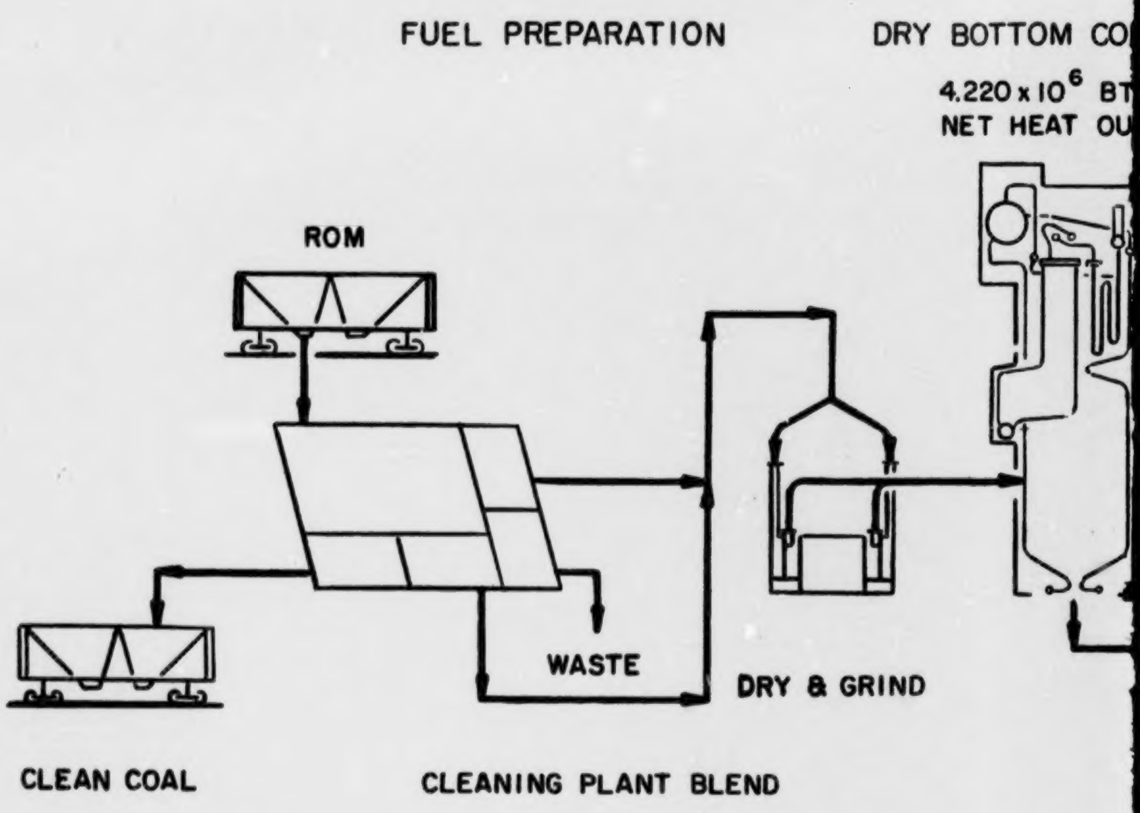
4. SUMMARY OF THE CASE D SYSTEM

(Refer to diagram following)

A run of mine bituminous coal containing 3 to 4.5% sulfur, washable to 1% sulfur, is assumed. The reject is treated for maximum yield of an HSC Fuel of the CASE C and D specification. The fuel is pulverized and fired to raise 3.5 million pounds per hour of subcritical steam which is fed to a turbogenerator of 500 MW capacity. The flue gas composition is identical to that of CASE C and is treated in an identical manner to yield 770 net tons per day of sulfur. Sulfur emission equals 0.8% of sulfur charged. Sulfur not emitted equals 16 pounds per million Btu of heat recovered in the CASE D System, or 2.3 pounds per million Btu of ROM coal charged to the cleaning plant.

CASE D is described in greater detail in Volume Two, Part XVI.

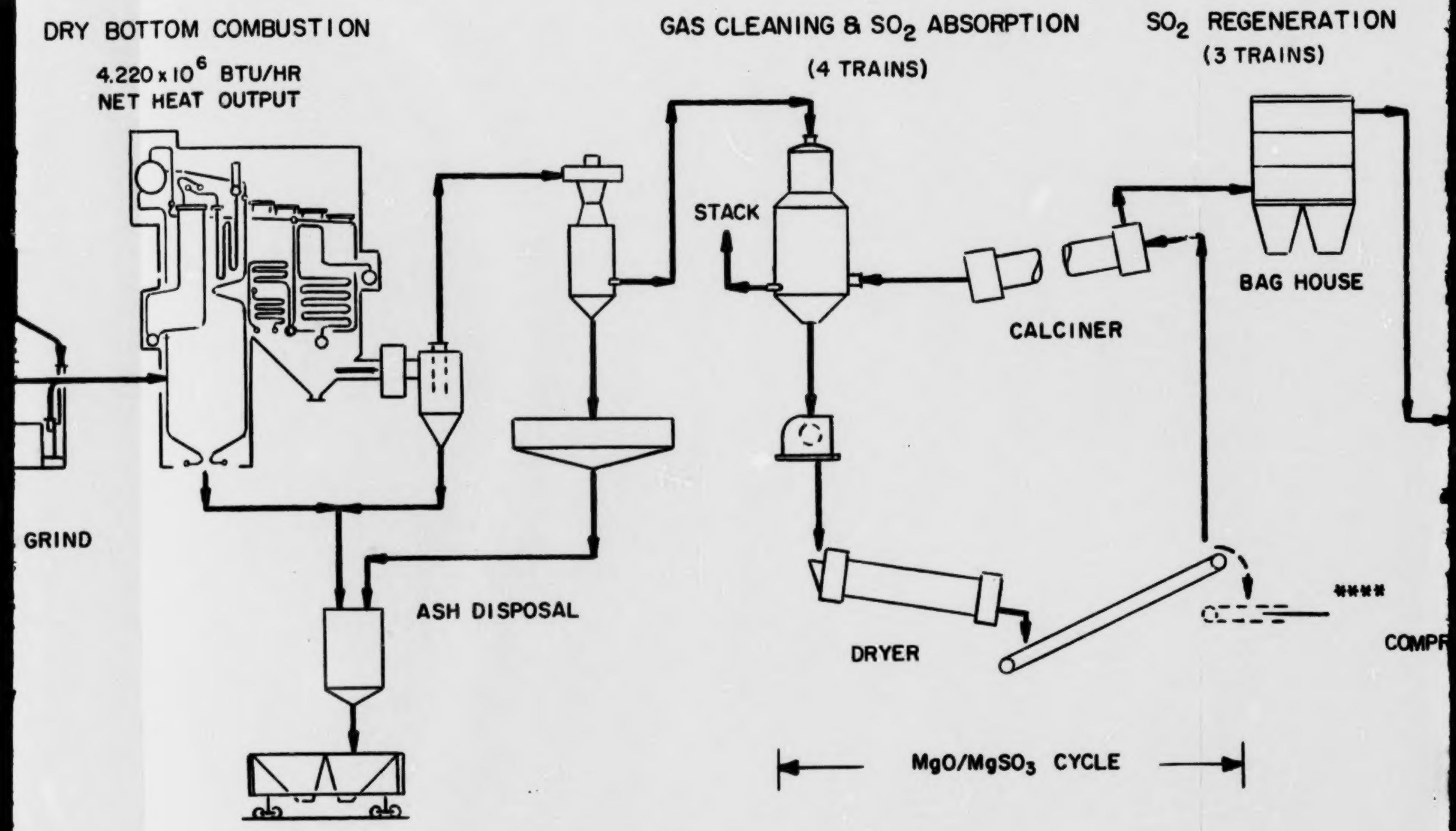
a


FUEL PREPARATION
DRY BOTTOM CO
4.220 x 10^6 BT
NET HEAT OU
ROM
WASTE
DRY & GRIND
CLEAN COAL
CLEANING PLANT BLEND

	RUN OF MINE	CLEAN COAL	HSC FUEL
IRON PYRITE			20.8
ORGANIC SULFUR			0.6
TOTAL SULFUR	3 TO 4.5		(11.7)
COAL COMBUSTIBLES			58.6
NON PYRITIC ASH			20.0
M TONS/YEAR			2280
TONS/HR: IRON PYRITE			59.3
COAL COMBUSTIBLES			338.0
NON PYRITIC ASH			57.0
SULFUR/COAL RATIO			0.2
BTU/LB			9000
BTU/HR			5,128 x 10^6

* 6% MOISTURE ASSUMED IN HSC FUEL AS FIRED

B


DRY BOTTOM COMBUSTION
4.220 x 10^6 BTU/HR
NET HEAT OUTPUT
GAS CLEANING & SO2 ABSORPTION
(4 TRAINS)
SO2 REGENERATION
(3 TRAINS)
STACK
BAG HOUSE
CALCINER
GRIND
ASH DISPOSAL
DRYER
****
COMPR
MgO/MgSO3 CYCLE

HSC FUEL
20.8
0.6
(11.7)
58.6
20.0
2280
59.3
338.0
57.0
0.2
9000
5,128 x 10^6

FLUE GAS VOL % DRY BASIS**		ASH COMPOSITION		CALCINER EMISSION	
SO_2	0.1	%FE	29	SO_2	15.9
SO_3	0.05	TONS/HR	96.5		
O_2	4.3 ***				
CO_2	15.4			CO_2	13.6
N_2	79.15			N_2	70.5

TEMP °F	500
MACFM	2,050
MSCFM	1,110

** NOMINAL MOISTURE CONTENT 4%

*** 25% EXCESS COMBUSTION AIR

**** MAGNESIUM SALTS MAY BE SHIPPED BEFORE PROCESSIN

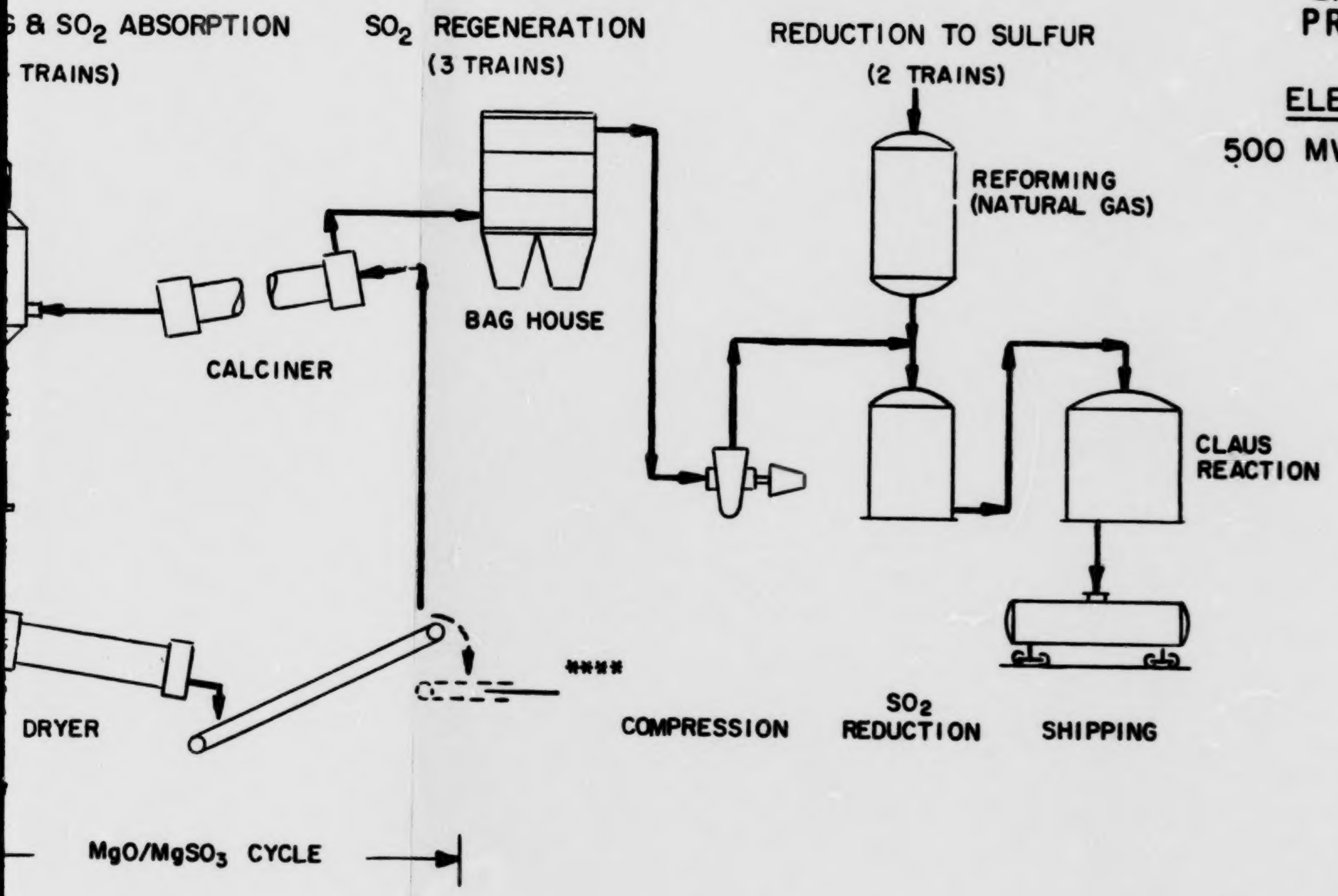
C

CASE D

ENERGY PRODUCT

ELECTRICITY

500 MW CAPACITY


S & SO_2 ABSORPTION
TRAINS)
SO_2 REGENERATION
(3 TRAINS)
REDUCTION TO SULFUR
(2 TRAINS)
BAG HOUSE
REFORMING
(NATURAL GAS)
CALCINER
CLAUS
REACTION
DRYER
****
COMPRESSION
SO_2
REDUCTION
SHIPPING
$MgO/MgSO_3$ CYCLE

CHEMICAL PRODUCT

COMMERCIAL GRADE

SULFUR

770 TONS/DAY.

ASH COMPOSITION		CALCINER EMISSION	
%FE	29	SO_2	15.9
TONS/HR	96.5		
		CO_2	13.6
		N_2	70.5

**** MAGNESIUM SALTS MAY BE SHIPPED BEFORE PROCESSING

ON AIR

5. SUMMARY OF THE CASE E SYSTEM

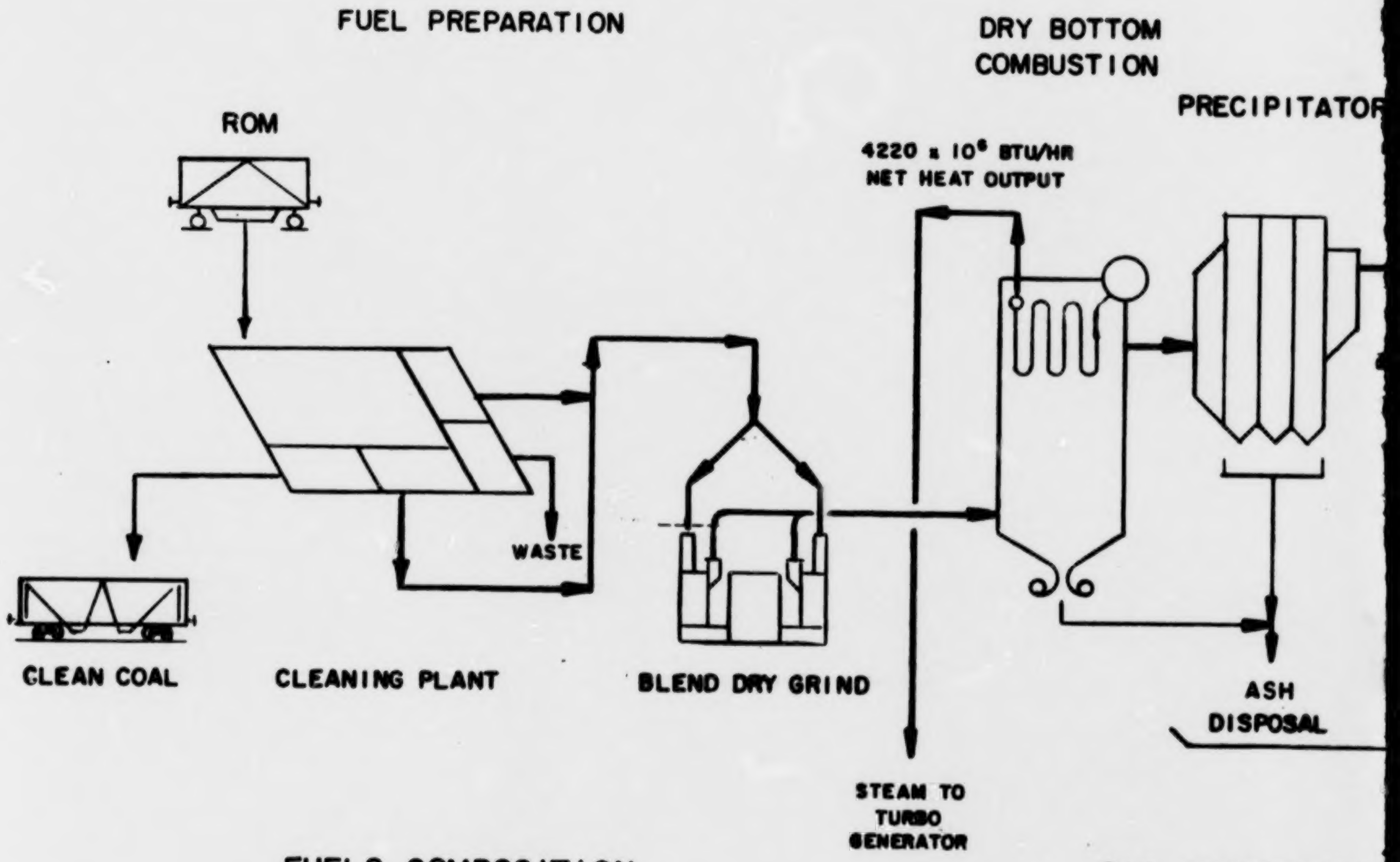
(Refer to diagram following)

A run of mine bituminous coal containing 3 to 4.5% sulfur, washable to 1% sulfur, is assumed. The reject is treated for maximum yield of an HSC Fuel of the CASE E specification. This fuel is pulverized and fired to raise 3.5 million pounds per hour of sub-critical steam which is fed to a turbo-generator of 500 MW capacity. The flue gas containing 0.7% sulfur dioxide and 4.2% oxygen is treated by the Monsanto Cat-Ox process in a manner similar to CASE B, but in multiple trains, to yield 1,420 net tons (100% basis) of 77/80% sulfuric acid per day. Sulfur emission equals 1.5% sulfur charged. Sulfur not emitted equals 10 pounds per million Btu recovered in the CASE E System, or 2.3 pounds per million Btu of ROM coal charged to the cleaning plant.

CASE E is described in greater detail in Volume Two, Part XVII.

a

HIGH SULFUR COMBUSTOR STU


FUEL PREPARATION
DRY BOTTOM COMBUSTION
PRECIPITATOR
ROM
4220 x 10⁶ BTU/HR NET HEAT OUTPUT
WASTE
CLEAN COAL
CLEANING PLANT
BLEND DRY GRIND
ASH DISPOSAL
STEAM TO TURBO GENERATOR

FUELS COMPOSITION
WT % DRY BASIS*

	RUN OF MINE	CLEAN COAL	HSC FUEL
IRON PYRITE			13.2
ORGANIC SULFUR			0.9
TOTAL SULFUR	3 TO 4.5		(7.9)
COAL COMBUSTIBLES			65.9
NON PYRITIC ASH			20.0
M TONS/YEAR			2133
TONS/HR: FeS_2			35.2
COAL			178.0
N-P ASH			53.3
SULFUR/COAL RATIO			0.12
BTU/LB			9660
BTU/HR			5150×10^6

*15% MOISTURE ASSUMED IN HSC FUEL AS FIRED

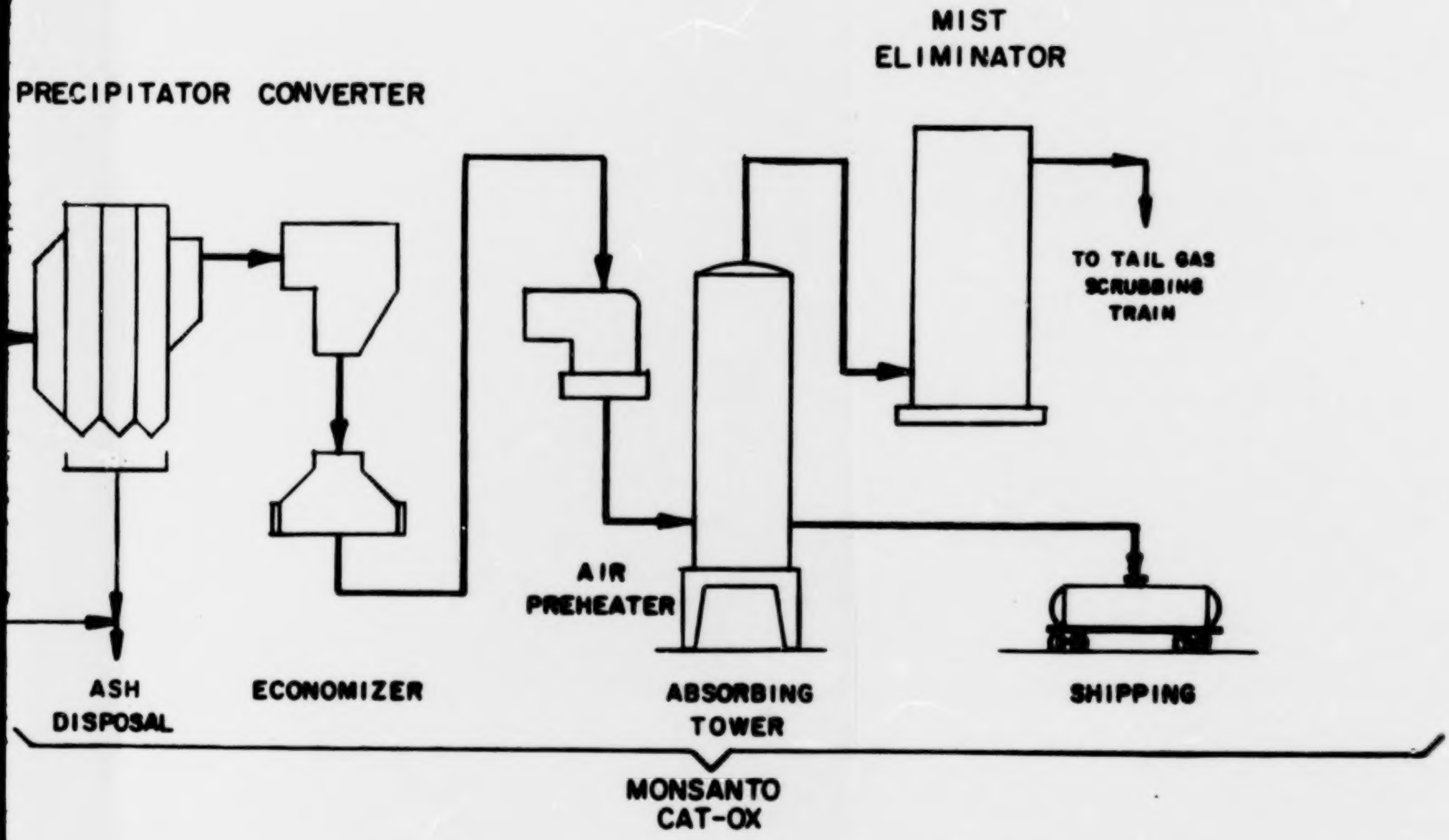
GAS COMPOSITI
VOL % DRY BAS

	EXIT FURNACE
SO_2	0.7
SO_3	0.02
O_2	4.2
CO_2	15.9
N_2	79.18
H_2O	**
TEMP °F	850
M ACFM	3080
M SCFM	1160
ASH	
TON/HR	76.7
% Fe	21

USTOR STUDY

CASE E
ENERGY
PRODUCT
ELECTRICITY
500 MW CAPACITY

B


MIST ELIMINATOR
PRECIPITATOR
CONVERTER
TO TAIL GAS SCRUBBING TRAIN
AIR PREHEATER
ASH DISPOSAL
ECONOMIZER
ABSORBING TOWER
SHIPPING
MONSANTO CAT-OX

COMPOSITION
% DRY BASIS

	EXIT FURNACE
SO_2	0.7
SO_3	0.02
O_2	4.2
CO_2	15.9
N_2	79.18
H_2O	**
TEMP °F	850
M ACFM	3080
M SCFM	1160
ASH	
TON/HR	76.7
% FE	21

	NO. OF UNITS
PRECIPITATORS	6
CONVERTERS	12
ECONOMIZERS	6
AIR HEATERS	3
ABSORBING TOWERS	3
MIST ELIMINATORS	3

CHEMICAL PRODUCT

COMMERCIAL
GRADE 77/80%
SULFURIC ACID

1420 TONS/DAY
100% BASIS

** NOMINAL MOISTURE CONTENT 4%

VIII
ECONOMIC EVALUATION

1. METHOD

The High Sulfur Combustor is conceived as a salvage system to improve the economics of coal cleaning by the sale of energy and sulfur value extracted from the reject fraction. Sales income, less all costs of extraction and other expense, is a basis for valuation of the HSC Fuels at the point of use of such fuels. The "point of use" would normally be input to the pulverizing mills but these fuels may have been milled to p.c. size prior to cleaning (see p. II-3). If so, the fuel would be fired directly in modified equipment. In any event, some costs will have been incurred upstream of this point of use during preparation of the HSC Fuel from the reject material, and for transportation, storage and handling. Note, however, that these costs are partly or wholly offset by the alternative cost of dumping all the reject material.

1.1 Valuation Formula

The HSC Fuel valuation formula adopted for the purpose of this study can be summarized as follows:

HSC Fuel Value = SV - SE - EC - G&T - PTH + D

Where SV = Sales Value
SE = Selling and Commercial Expense
EC = Extraction Cost
G&T = General Expense and Income Tax
PTH = HSC Fuel Preparation, Transport and Handling ("Precombustion" costs)
D = Cost of Dumping Refuse

1.2 Precombustion Costs vs. Dumping

Of the several terms in this valuation formula the last two are least amenable to estimation out of context. The composition of the coal to be cleaned, the characteristics of the operating locality, proximity of the combustor to the cleaning plant, are all cost determinants of importance. While the cost of coal washing is not a term in the formula it is nevertheless suggestive: 42¢ per ton of clean coal (at 90% yield) Ref (F), up to 76¢ Ref (KKb). The cost of preparing the HSC Fuel and handling into the pulverizers will vary at least as widely. One investigator has estimated this cost at 9¢ per ton of shipped (i.e. cleaned) coal, or 81¢ per ton of HSC Fuel of sulfur content in the range 7 to 9% Ref (KKb). On the other hand, where on-site dumping continues to be possible, the cost of it is estimated all the way from $1.00 to $2.00 per ton of refuse, depending mainly on the proximity and rental value of dumping ground. In these circumstances the cost assumption for the purpose of this study is that it will be as cheap to prepare and deliver an HSC Fuel as to dump an equal tonnage of refuse: that is to say, minus PTH plus D equals zero in the valuation formula above.

2. COSTS, PRICES AND ESCALATION

2.1 Capital Costs

Since this study is unavoidably generalized and speculative in several aspects the economic estimates do not reflect any very long-term escalation factors. Construction labor, plant equipment and material, are priced at the high end of the range expected in 1971. Prices of major equipment items elicited from vendors in late 1970 were for 12 to 18 month delivery. On balance the capital estimates may thus be reasonably realistic for 1972. During and beyond this time it is noted that capital cost escalations as great as 10% annually may be experienced in the construction of new thermo-electric generating capacity. Capital cost escalation for chemical plant is expected to be less extreme -- of the order of 6 to 8% per year.

The estimated capital costs of the HSC Systems are summarized below:

TABLE VIII-1

ESTIMATED CAPITAL COST

	CASE A $ M	CASE B $ M	CASE C $ M	CASE D $ M	CASE E $ M
Combustor/Boiler	5, 110	4, 210	4, 310	15, 760	11, 760
Electric Equipment			11, 200	56, 200	56, 200
Sulfur Recovery	11, 300	12, 500	10, 900	44, 000	45, 700
Off-Sites	1, 550	1, 550	2, 550	10, 100	10, 500
Land	250	250	250	1, 000	1, 000
Fixed Capital	18, 210	18, 510	29, 210	128, 060	125, 160
Working Capital	130	90	60	460	410
Total Capital	18, 340	18, 600	29, 270	128, 520	125, 570

2.2 Costs of Operation or "Extraction"

2.2.1 The utilities group of cost elements is largely dependent on the cost of various energy forms. Electricity is entered at 7.5 mills/Kwh, the same as estimated sales value discussed in a later paragraph. The designs of the sulfur recovery trains of CASE C and D require low-sulfur oil or gas for magnesium sulfite drying, oil or gas that may be high-sulfur for calcination, and a light hydrocarbon fuel for reforming to effect SO_2 reduction. CASE B and E require gaseous fuel for acid concentrating. Heat energy for these various uses is costed at 40¢ per million Btu in the estimates. By comparison the average delivered cost of steam coal reported by the American Coal Association for the six-state area was 23.8¢ per million Btu in 1969 -- probably about 25¢ in late 1970. Steam where needed for process is entered at the estimated sales value of $1.30 per ton. No credit is taken or given for condensate return except for the high recovery of condensate from the generator drives.

2.2.2 Operating labor and supervision are entered at the high end of rates expected in 1971.

2.2.3 Maintenance cost estimates are based on FPC studies of furnace and electric equipment (adjusted upward for near future conditions on the advice of Ebasco Services, Inc. and Foster Wheeler Corporation). Maintenance of the sulfur recovery equipment is entered at 4% annually of the depreciable capital cost.

2.2.4 Local taxes and insurance, as operating cost elements, will continue to increase in dollar amounts as a function of fixed and working capital, and probably at increasing rates, especially local taxes. The cost of money is given effect in the operating estimates at 9% annually on undepreciated investment as the resultant of estimates as follows:

Internal Capital	20% @ 7%	=	1.40
Debt Capital	60% @ 7%	=	4.50
Equity Capital	20% @ 15%	=	3.00
Average			8.90%

2.2.5 Depreciation is figured, straightline, at 3.6% (28 year life) on furnace and electric equipment, and 9.1% (11 year life) on the flue gas treating trains; that is the chemical equipment. These are "guideline" rates from IRS Publication #456. Capital (and capital related) charges to operations are thus in the range 13 to 16% of total investment annually. The variation within this range results from the two depreciation rates.

2.2.6 Product Costing Basis

In Tables VIII-2 through VIII-6 the estimated cost of operation or extraction (that is "EC" in the valuation formula) of each of the five systems is set up in the conventional manner except that no charge is entered for the HSC Fuel. Additionally, since each system yields two products, energy and sulfur value, an effort is made to cost each product separately. For this purpose the combustor/generator operation, that is, everything up stream of the cooled flue gas, is charged to energy output, and the SO_2 recovery train, downstream of the cooled flue gas, is charged to recovered sulfur value. CASE B and E systems recycle some heat value from the acid converters back to the boilers. An adjustment is entered for this.

TABLE VIII-2

CASE A SYSTEM

ESTIMATED OPERATING COST

	Unit	Unit Value	STEAM Units Per Year	STEAM Annual Cost	SULFURIC ACID Units Per Year	SULFURIC ACID Annual Cost
Production						
Energy	N. T.		2,000,000			
Sulfur Value	N. T.				610,000	
Utilities, etc.						
Electricity	MWH	$7.50	6,080	M$ 46	24,630	M$ 185
Cooling Water	MMgal.	8.00			1,400	11
Boiler Feed Water	Mgal.	0.60	475	285		
Process Water	Mgal.	0.40			9	40
Steam	N. T.	1.30			389,000	506
Chemicals						413
Waste Disposal	N. T.	2.00	390,000	780	120,000	240
Labor						
Supervision	Manhour	8.00	4,380	35	4,380	35
Operators and Helpers	Manhour	5.50	15,000	83	11,000	61
Maintenance				140		450
Plant General Expense						
Factory Overhead	80% of Direct Labor			66		49
Taxes and Insurance	2.5% of Total Capital			152		306
Total Before Depreciation and Interest				M $1,587		M $2,296
Depreciation						
Guideline Life	11 Years	9.1%		664		970
Interest (Average)						
Depreciable Capital		4.5%		275		533
Non-Depreciable Capital		9 %		17		17
Cost Before HSC Fuel Charge (Rounded)				M $2,540		M $3,820

TABLE VIII-3

CASE B SYSTEM

ESTIMATED OPERATING COST

	Unit	Unit Value	STEAM Units Per Year	STEAM Annual Cost	SULFURIC ACID Units Per Year	SULFURIC ACID Annual Cost
Production						
Energy	N. T.		2, 000, 000			
Sulfur Value	N. T.				208, 000	
Utilities, etc.						
Electricity	MWH	$7. 50	5, 000	M$ 38	115, 400	M$ 870
Cooling Water	MMgal.	8. 00			2, 854	23
Boiler Feed Water	Mgal.	0. 60	479, 000	287		
Fuel Gas	MMBtu	0. 40			294, 000	118
Catalyst						25
Chemicals						194
Waste Disposal	N. T.	2. 00	140, 000	280	20, 350	41
Labor						
Supervision	Manhour	8. 00	4, 380	35	4, 380	35
Operators and Helpers	Manhour	5. 50	15, 000	83	11, 000	61
Maintenance				140		464
Plant General Expense						
Factory Overhead	80% of Direct Labor			66		49
Taxes and Insurance	2. 5% of Total Capital			113		351
Total Before Depreciation and Interest				M $1, 042		$2, 231
Depreciation						
Guideline Life	11 Years	9. 1%		553		1, 109
Interest (Average)						
Depreciable Capital		4. 5%		274		548
Non-Depreciable Capital		9 %		10		21
Cost Before HSC Fuel Charge (Rounded)				M $1, 880		M $3, 910

TABLE VIII-4

CASE C SYSTEM
ESTIMATED OPERATING COST

			ELECTRICITY		SULFUR	
	Unit	Unit Value	Units Per Year	Annual Cost	Units Per Year	Annual Cost
Production						
Energy	MWH		400,000			
Sulfur Value	L. T.				30,910	
Utilities, etc.						
Electricity	MWH	$7.50	3,360	M$ 25	37,360	M$ 280
Cooling Water	MMgal.	8.00	21,600	173	268	2
Boiler Feed Water	Mgal.	0.60	5	3		
Process Water	Mgal.	0.40			113,000	45
Fuel Gas	MMBtu	0.40			2,100,000	840
Chemicals						81
Waste Disposal	N. T.	2.00	110,000	220	25,400	51
Labor						
Supervision	Manhour	8.00	8,760	70	8,760	70
Operators and Helpers	Manhour	5.50	44,000	242	32,000	176
Maintenance				280		430
Plant General Expense						
Factory Overhead	80% of Direct Labor			194		141
Taxes and Insurance	2.5% of Total Capital			403		329
Total Before Depreciation and Interest				M$1,610		M$2,445
Depreciation						
Guideline Life	11 Years	9.1%				1,174
Guideline Life	28 Years	3.6%		578		
Interest (Average)						
Depreciable Capital		4.5%		723		581
Non-Depreciable Capital		9 %		153		126
Cost Before HSC Fuel Charge (Rounded)				M$3,060		M$4,330

TABLE VIII-5

CASE D SYSTEM

ESTIMATED OPERATING COST

			ELECTRICITY		SULFUR	
	Unit	Unit Value	Units Per Year	Annual Cost	Units Per Year	Annual Cost
Production						
Energy	MWH		4, 000, 000			
Sulfur Value	L. T.				233, 300	
Utilities, etc.				M		M
Electricity	MWH	$7. 50	24, 000	$ 180	282, 300	$ 2, 117
Cooling Water	MMgal.	4. 00	86, 400	346	4, 000	16
Boiler Feed Water	Mgal.	0. 60	50	30	6	4
Process Water	Mgal.	0. 40			852, 000	341
Fuel Gas	Btux10^9	400.			17, 830	7, 132
Chemicals				100		535
Waste Disposal	N. T.	2. 00	853, 800	1, 708	192, 000	384
Labor						
Supervision	Manhour	8. 00	8, 760	70	8, 760	70
Operators and Helpers	Manhour	5. 50	65, 000	360	48, 000	264
Maintenance				1, 080		1, 760
Plant General Expense						
Factory Overhead	80% of Direct Labor			288		211
Taxes and Insurance	2. 5% of Total Capital			1, 992		1, 221
Total Before Depreciation and Interest				M$6, 154		M$14, 055
Depreciation						
Guideline Life	11 Years	9. 1%				4, 353
Guideline Life	28 Years	3. 6%		2, 816		
Interest (Average)						
Depreciable Capital		4. 5%		3, 545		2, 173
Non-Depreciable Capital		9 %		81		50
Cost Before HSC Fuel Charge (Rounded)				M$12, 600		M$20, 630

TABLE VIII-6

CASE E SYSTEM

ESTIMATED OPERATING COST

			ELECTRICITY		SULFURIC ACID	
	Unit	Unit Value	Units Per Year	Annual Cost	Units Per Year	Annual Cost
Production						
Energy	MWH		4, 000, 000			
Sulfur Value	N. T.				473, 300	
Utilities, etc.						
Electricity	MWH	$7. 50	20, 000	M$ 150	222, 000	M$ 1, 665
Cooling Water	MMgal.	4. 00	86, 400	346	26, 500	106
Boiler Feed Water	Mgal.	0. 60	50	30		
Fuel	MMBtu	0. 40			730, 000	292
Chemicals	N. T.			100		505
Waste Disposal	N. T.	2. 00	614, 000	1, 228	492, 000	984
Labor						
Supervision	Manhour	8. 00	8, 760	70	8, 760	70
Operators and Helpers	Manhour	5. 50	65, 000	360	16, 000	880
Maintenance				1, 080		1, 668
Plant General Expense						
Factory Overhead	80% of Direct Labor			288		704
Taxes and Insurance	2. 5% of Total Capital			1, 883		1, 256
Total Before Depreciation and Interest				M$ 5, 535		M$ 8, 130
Depreciation						
Guideline Life	11 Years	9. 1%				4, 541
Guideline Life	28 Years	3. 6%		2, 673		
Interest (Average)						
Depreciable Capital		4. 5%		3, 342		2, 246
Non-Depreciable Capital		9. 0%		76		51
Cost Before HSC Fuel Charge (Rounded)				M$11, 630		M$14, 970

2.2.7 Pollution Control

The foregoing product costing basis is oversimplified to the extent that it leads, as expected, to an apparent low cost of energy extraction and a high cost of sulfur value recovery. But, of course, it does not follow that the energy alone could be produced and sold while avoiding the cost of sulfur value recovery, or at least the cost of pollution control, since in any event the SO_2 and particulate matter will have to be cleaned out of the combustion gases. A separate estimate of the cost of "pollution control" has therefore been developed for each system sufficient to reduce the fly ash and the SO_2 to wet, inert, solid material for dumping. The estimates reflect the use of venturi-type scrubbers circulating calcium oxide slurry to capture particulate matter virtually completely and reduce the final stack emission of SO_2 to less than 200 ppm. The estimated investment and annual operating cost applicable to the combustion gas of each system is as shown in Table VIII-7.

Note particularly that both investment and operating costs exhibited in Table VIII-7 are very high by comparison to pollution control costs on conventional coal-fired steam generators of similar sizes, even though flue gas volumes are not greatly different. Investment cost differences result principally from the multi-stage scrubbing required to reduce these high SO_2 concentrations to an acceptable level, and provide equipment for handling waste solids several times greater than usual. The higher operating costs are mainly a result of calcium oxide cost and waste disposal cost.

TABLE VIII-7

"POLLUTION CONTROL" AS ALTERNATE TO SULFUR VALUE RECOVERY

			CASE A		CASE B		CASE C		CASE D		CASE E	
Fixed Capital				M$ 7, 000		M$ 7, 000		M$ 5, 200		M$ 18, 000		M$ 16, 500
Working Capital				700		240		120		800		560
Total				7, 700		7, 240		5, 320		18, 800		17, 060
Operating Cost	**Unit**	**Price Per Unit**	**Used Per Year**	**Cost Per Year**	**Used Per Year**	**Cost Per Year**	**Used Per Year**	**Cost Per Year**	**Used Per Year**	**Cost Per Year**	**Used Per Year**	**Cost Per Year**
Lime (CaO)	Net Ton	15. 00	400, 000	M$ 6, 000	142, 000	M$ 2, 130	66, 000	M$ 990	466, 000	M$ 7, 000	320, 000	M$ 4, 000
Water	M. Gals	0. 40	72, 000	29	100, 000	40	144, 000	58	527, 000	211	523, 000	209
Electricity	MWH	7. 50	15, 200	114	25, 600	192	23, 800	179	203, 000	1, 523	203, 000	1, 523
Waste Disposal	Net Tons	2. 00	852, 000	1, 704	325, 000	650	135, 000	270	940, 000	1, 880	286, 000	572
Operators	Man Hours	5. 50	8, 000	44	8, 000	44	8, 000	44	16, 000	88	16, 000	88
Maintenance	4% of Fixed Capital			280		280		208		720		660
Factory Overhead	80% of Direct Labor			35		35		35		70		70
Taxes and Insurance	2. 5% of Total Capital			193		181		133		470		427
Depreciation	9. 1% of Fixed Capital			637		637		473		1, 638		1, 502
Interest	4. 5% of Total Capital			347		326		239		846		768
Total (Rounded)				M$ 9, 380		M$ 4, 520		M$ 2, 630		M$ 14, 450		M$ 10, 620

N. B. The pollution control duty reflected in these figures is much more severe than for conventional steam generators. See preceding page.

2.3 Sales Value of Products

Unit prices of HSC products that are given effect herein, have been determined by a canvass of buyers and vendors now in business in the six-state area.

2.3.1 Steam of 825° F and 900 psig at the rate of 500,000 lbs/hr is a product of CASE A and CASE B to be consumed at the producing site or close thereto. While low pressure steam is sold as a public utility in cities, for example Chicago, Rockford, Indianapolis, to schools, hospitals, small industry and the like, the steam is usually a bleed product from generator drive-turbines, and the cost of distribution is too great for any but densely populated areas. Most of the steam now sold in this way is distributed through long existing pipe facilities. Prices appear to vary from $1.20 to $3.00 per ton depending on quantity and load factor.

The realistic market possibility is an existing steam-electric plant or a large industrial consumer, more likely the latter. Chemical and metallurgical works are typically users of steam in substantial quantities. Questions concerning steam use and value were made a feature of the field inquiry into sulfur and sulfuric acid manufacture and use. Much of the traffic in industrial steam involves payment with off-gases, petroleum residues and the like, but a consensus favors the neighborhood of $1.30 per ton for near future real value. This figure is used.

2.3.2 Electricity is priced for sale at $7.50 per megawatt hour, (7.5 mills/kwh) as a median between a generally prevailing level of $9.00 for peak power and $6.00 for off-peak. Both values will move upward with increase in fuel and other costs, but certainly less fast than the increases expected in new plant costs. A negotiated sale of power in the six-state area would reflect the demand pattern of the expected resale market. Substantial differences prevail between urban and rural, north and south.

2.3.3 Sulfur was shown in Figure III-3 to be consumed at a rate approaching 10 million long tons per year in the U.S. Of this amount about 7 million long tons was produced from U.S. Frasch process sources, down about 5% from 1969. At the same time Canadian production increased about 0.6 million long tons over 1969 to a 1970 total of 4.4 million long tons. A pattern of declining U.S. and Mexican production of newly mined sulfur and increasing recovery of by-product sulfur in Canada and in the U.S. is expected to continue indefinitely. In 1970 for the first time total world output of recovered sulfur exceeded Frasch process output. But whatever the source of elementary sulfur, more than 90% of production is converted to sulfuric acid and used as such; and the recovery of sulfuric acid directly from waste acids and waste SO_2 is also in an up-trend that will continue indefinitely. These trends, now visible in initial stages, are in large part a consequence of the effort to control emission of sulfur oxides from furnaces, smelters and the like. As this effort continues and gains in scope, a very extensive change will occur in the historical basis for sulfuric acid production. With the exception of elementary sulfur recovered during the preparation of sour natural gas for market, Frasch process sulfur has been the least costly source; an extraction cost of $7 per long ton has been estimated for sulfur from the best Gulf Coast mines. Sulfur from sour gas was offered at $C9.33 per long ton f.o.b. Alberta, Canada in late 1970. The actual cost of sulfur recovery from sour gas is usually less than this. With this background recent delivered prices of sulfur were as follows:

	Per LT Delivered
Gulf Region	$25 - $27
Tampa Area	27 - 29
Atlantic Seaboard	29 - 35
No. Central States	24 - Plus

Sulfur delivered to the six-state area of special interest in this study (Pennsylvania, Ohio, West Virginia, Kentucky, Indiana and Illinois) originates mostly in Alberta for the north and west of this region and Gulf Coast sources for the south and east. Late 1970 delivered prices of Canadian sulfur were:

	Per LT Delivered
Chicago, Ill. Madison, Wisc. Hammond, Ind.	$24.51
Detroit, Mich. St. Louis, Mo.	$26.75
Cleveland, Ohio Cincinnati, Ohio	$28.09

All the foregoing were based on $U.S. 9.50 per LT f.o.b. Alberta, Canada. Sulfur shipments to such points have increased steadily from 568,000 long tons in 1964 to 907,000 long tons in 1969. Delivered prices of sulfur in the south and east of this region reflect a Gulf Coast price of about $25 per long ton plus $3 to $5 delivery cost. Deliveries are mostly as liquid by barge to producers' barge terminals and locally by tank truck.

The six-state region may consume upwards of 1,000,000 long tons per year of sulfur in acid making. See Table VIII-1. A sulfur recovery unit such as CASE D, capable of producing 233,000 long tons per year can account by itself for one fifth or more of this requirement. A new source of supply of such magnitude will depress sulfur price in these areas even against the general price inflation looked for in the next few years. Note, however, that Gulf Coast sulfur vendors have retreated rather than compete in a drastic price war with Canadian imports. The sticking point, for example at Cincinnati, appears now to be about $28 per long ton corresponding to a Gulf Coast price of $23. Continued pressure may result in some reduction in this level but fact and opinion combine to indicate that domestic price will not go below $20 f.o.b. Gulf port. This view is confirmed by private communication with two major sulfur vendors. Accordingly, the sulfur sales value given effect in this study is $25 per long ton delivered to points in the six-state area.

TABLE VIII-8

SULFURIC ACID
ESTIMATED PRODUCTION AND SHIPMENTS
1968

State	Total Capacity, TPY (3)	Production, TPY	Shipment, TPY
Pennsylvania	920, 200	916, 279(1)	850, 725(1)
Ohio	753, 500	629, 229(1)	519, 931(1)
Illinois	2, 178, 000	1, 724, 854(1)	812, 633(1)
Indiana	672, 000	571, 000(2)	485, 500(2)
Kentucky	254, 000	216, 000(2)	183, 500(2)
West Virginia	135, 000	115, 000(2)	98, 500(2)
Totals	4, 912, 700	4, 172, 362	2, 950, 289
Estimated Capacity Based on Sulfur	3, 619, 000		
Estimated Production From Sulfur		3, 100, 000	
Estimated Sulfur Consumption		1, 000, 000	

(1) Bureau of the Census, Current Industrial Reports, Sulfuric Acid, 1968, M28A (68)-13 Supplement 1.

(2) CHEMICO Estimates: Production at 85% of capacity; shipments at 85% of production.

(3) Net Tons per Year

2.3.4 Sulfuric Acid

The versatility and low cost of this product have made its usefulness almost coextensive with all industrial activity; the long term growth of consumption has substantially exceeded the growth of GNP. The six-state area is not a huge consumer of sulfuric acid, however, since more than 40% of U.S. production (See Table VIII-9) is used in phosphate fertilizer manufacture which is practiced only in Illinois of the six-states. The outlook for growth of consumption in this area during the 1970's is now estimated at 3 to 5% annually even though some important uses are declining and will probably continue to decline.

2.3.4.1 The major end-uses of sulfuric acid in the six-states are as follows:

Phosphate rock processing for fertilizer may consume as much as 70% of the sulfuric acid made in Illinois which is, by reason of this single demand, the big acid producer among the six-states. This use thus accounts for 1.2 million tons per year; but is currently declining with the general decline in fertilizer use in recent years. This trend is not expected to continue for long.

All of the rock processors in Illinois are also manufacturers of sulfuric acid.

Ammonium sulfate, is made from coke-oven ammonia in Pennsylvania, West Virginia, Indiana and Ohio, and is made also in the area as a byproduct of caprolactam manufacture. A big drop in sulfuric acid price might slow down the declining output trend in this product, demand for which is not brisk since it is mainly a source of nitrogen for fertilizer and rather dilute for this purpose. Even so ammonium sulfate may account for 400, 000 tons per year of sulfuric acid consumption in the six-states.

Coke-oven ammonia is treated alternatively with phosphoric acid to make diammonium phosphate, a concentrated and popular fertilizer. This practice may grow and ultimately increase demand for coke-oven ammonia for fertilizer. If so the use of sulfuric acid will be sustained since the tonnage required is about the same for either the phosphate or sulfate salt of ammonia.

The use of sulfuric acid for iron and steel pickling is still large in Pennsylvania, Indiana and Ohio although hydrochloric acid is invading this domain and probably now accounts for 30% of total pickling acid. Hydrochloric acid, initially used as a substitute during the period of rapidly increasing sulfuric acid price (1965-1969), is now claimed by many to do a better job. Probably the expected lower, future cost levels of sulfuric acid will slow down or reverse the trend to hydrochloric. Sulfuric acid consumption for pickling is estimated at 600,000 tons per year in the six-states.

Petroleum refining and Petrochemistry consume sulfuric acid as a catalyst in production of alkylates (antiknock compounds) for gasoline, and in sulfonation of lube oils. Alkylation spent acid is usually reworked, but net sulfuric acid consumption for this purpose should increase as the use of alkylated lead compounds gives way to growing pressure for pollution abatement. Sulfuric acid is also used to make hydrofluoric acid which, among many uses, is increasing in demand for manufacture of motor fuel alkylates. The outlook for growth of hydrofluoric acid manufacture is dimmed by the possibilities for increased recovery of hydrofluoric acid from phosphate rock processing, also a pollution abatement measure.

Sulfonated and sulfated surfactants and alcohols derived from petroleum fractions consume sulfuric acid importantly in the six-state area. The surfactants are the active components in wetting agents and detergents and have good growth potential among the biodegradable types.

The foregoing uses in organic chemistry probably account for 10% of acid consumption in the six-states, say 420,000 tons per year.

The major end-uses listed above appear to account for 2.6 million tons of sulfuric acid annually in the six-states, about 60% of 1968 production. A large number of uses are to be credited with the balance of consumption, of which alum, dyes and intermediates, and explosives manufacture, are significant in this area.

2.3.4.2 Sulfuric acid list prices in the United States increased gradually from $16/17 per net ton during the years 1930-1948, increased steadily thereafter to $23 that prevailed during the period 1953 to 1964, and increased rapidly to $28 per net ton in 1968. During all this time the acid prices rather faithfully paralleled the trend of sulfur prices. Discounts from list prices were always available to large volume buyers.

Sulfuric acid prices in the four major producing states of the six-state area tended to run above the national average as indicated below:

TABLE VIII-8

SULFURIC ACID
(Average Price of Shipments per net ton FOB)

From Plants in:	1964	1965	1966	1967	1968
Pennsylvania	$17.72	$17.92	$18.11	$20.08	$22.99
Ohio	20.26	22.31	22.89	24.77	27.99
Indiana	---	---	20.14	---	---
Illinois	19.00	19.85	18.92	19.34	20.56
U.S. Average	16.72	16.76	17.39	18.68	21.33

A late 1970 field investigation indicated that average price for medium to large volume contract purchases in markets such as Pittsburgh, Cleveland, Chicago and St. Louis is $25 to $28 per net ton, freight equalized. This is roughly equivalent to $22 to $26 FOB production point. Smaller volume sales are priced at $33 to $34 per net ton delivered.

The future price trend is certainly downward from the current level with ever-stiffer competition from by-product sources.

Perhaps most immediately significant to the northern portion of the six-state area (where most of the sulfuric acid is used) is the huge acid recovery operation planned for the nickel refineries to be built in Ontario. By 1975 as much as 5 million tons per year of acid capacity may be installed — an amount of acid capacity about twice the present level of Canadian consumption. Since freight costs *seriously limit* the quantities that can be shipped overseas, it is assumed that one or more million tons per year will be offered to nearby consumers in the U.S. at very low prices, especially to consumers accessible by water. Developments in pollution control, such as "central processing" of magnesium salts (Ref RR) carrying SO_2 recovered from combustion gases of fossil-fuel burning utility stations, more complete recovery of sulfide ore smelter gases, and the like, promise to add big new sources of sulfuric acid within the U.S. It is possible but believed not probable that competition from byproduct sources will drive sulfuric acid price well below the level at which production from Frasch process sulfur can profitably continue. This level for the six-state area is estimated as follows:

> Sulfuric acid requires on the average 0.31 long tons of sulfur per net ton of 100% acid. If, as argued in VII-2.3.3, Frasch process sulfur will not be priced below $25 per long ton delivered, the minimum FOB price of acid from large sulfur based plants is indicated to be --

	Net Ton
Sulfur @ $25 x 0.31	$7.75
Conversion Cost	2.00
Factory Cost	$9.75
Overhead and Profit	3.00
Minimum Price Per Net Ton 100% H_2SO_4	$12.75

This price is taken to be a conservative minimum since it is believed that Frasch Process sulfur or elementary sulfur from other sources will not be offered, in the six-state area, significantly below $25 per long ton. The market ought then to absorb by-product acid at this price or thereabouts. Normal growth of existing uses for sulfuric acid, and new uses stimulated by so low a price, will help. This price, $12.75, is given effect in the income estimates of Table VIII-9.

TABLE VIII-9

SULFURIC ACID USES IN THE UNITED STATES

'000 TONS

100% BASIS

	1964	1965	1966	1968
Phosphate Fertilizer	7,753	9,275	11,675	12,676
$(NH_4)_2SO_4$	1,759	1,967	1,885	1,815
TiO_2	1,975	1,900	1,740	1,230
Petroleum-Alkylate	1,375	1,400	1,435	1,332
-Other	1,205	1,260	1,335	1,400
Iron and Steel Pickling	1,190	1,075	1,000	794
HF	630	710	785	675
Rayon and Cello	885	895	836	772
Alcohols	631	665	700	706
Alum	574	580	592	643
Detergents	500	510	530	427
Copper and Uranium	576	567	550	923
Dyes and Intermediates	312	330	350	392
Explosives	118	100	198	900
Other Chemicals(1)	2,951	3,099	3,389	1,362
Exports	14	7	---	12
Miscellaneous(2)	1,302	1,360	1,500	2,746
	23,750	25,700	28,500	28,805

(1) Includes

	1965	1968
Chlorine Dioxide	120	211
Tall Oil	125	93
Pesticides	190	75
HCl	103	188
Borates	81	71
Chrome Chemicals	85	89
Chlorine Drying	100	123

(2) Includes

	1965	1968
Non-Ferrous Metals	275	316
Water Treatment	170	202
Batteries	140	128

3. OPERATING ECONOMICS

In Table VIII - 10 Income from Sales reflects the unit prices discussed in VIII - 2. Sulfuric acid recovered in CASE B and CASE E is of 93% concentration and entered at the same sales value as the 98% acid of CASE A. Both capital and operating cost estimates of CASE B and E include provision for concentration, in gas-fired drum-concentrators, of the Cat-Ox 77/80% acid (see pages VII-8 and VII-9) to 93%.

Selling and Commercial Expense are entered at a nominal 1% of sales value for the energy products on the theory that long term sales contracts would have been negotiated before location of the facility. Selling the sulfur products may be more continuously difficult - 5% of sales value is provided in the estimate.

"Operating Cost" carried forward from Tables VIII-2/6, includes interest but no charge for the HSC Fuel:

General & Administrative is entered at 2% of sales value.

The "Adjustment for Pollution Control (from Table VIII-7) is entered as a charge to energy recovery and a credit to sulfur value recovery. In CASE A and CASE B the credit is larger than the whole cost of sulfuric acid recovery. Note in CASE A, especially, the elimination of SO_2 from the flue gas for dumping as an inert solid is twice as costly as recovery of sulfuric acid for sale. Note also that the adjustment serves only the purpose of shifting the burden of "pollution control" from sulfur recovery to energy extraction in each CASE without affecting combined income, cost and gain.

TABLE VIII-10

INCOME FROM SALES, COSTS OF OPERATION AND PRODUCTS GAIN OR LOSS FROM OPERATION

ANNUAL BASIS

	Unit	Unit Value	Annual Quantity	CASE A		CASE B		CASE C		CASE D		CASE E	
				Steam	Sulfuric Acid	Steam	Sulfuric Acid	Electricity	Sulfur	Electricity	Sulfur	Electricity	Sulfuric Acid
Income From Sales													
Steam	Net Ton	$ 1.30	2,000,000	M$ 2,600		M$2,600							
Electricity	MWH	7.50	400,000					M$3,000					
			4,000,000							M$30,000		M$30,000	
Sulfur	Long Ton	25.00	30,910						M$ 770				
			233,300								M$ 5,830		
Sulfuric Acid (100%)	Net Ton	12.75	610,000		M$7,780								
			208,000				M$2,650						
			473,300										M$ 6,040
Less: Selling & Commercial Expense @ 1% of Sales Value				26		26		30		300		300	
@ 5% of Sales Value					389		132		39		292		302
Operating Cost (From Tables VIII-2 through 6)				2,540	3,820	1,880	3,910	3,060	4,330	12,600	20,630	11,630	14,970
General & Administrative @ 2% of Sales				52	156	52	53	60	15	600	117	600	121
Total Cost Before U.S. Income Tax (Rounded)				M$ 2,620	M$4,370	M$1,960	M$4,100	M$3,150	M$ 4,380	M$13,500	M$21,000	M$12,500	M$15,400
Adjustment for "Pollution Control" (Table VIII-7)													
Add				9,380		4,520		2,630		14,450		10,620	
Deduct					9,380		4,520		2,630		14,450		10,620
Adjusted Cost of Energy Extraction				M$12,000		M$6,480		M$5,780		M$28,000		M$23,100	
Unit Cost of Energy Products:													
Per Net Ton of Steam				$ 6.00		$ 3.20							
Per Megawatt Hour								$ 14.50		$ 7.00		$ 5.80	
Incremental Cost of Sulfur Value Recovery					(M$5,010)		(M$ 420)		M$ 1,750		M$ 6,550		M$ 4,780
Unit Cost of Sulfur Products:													
Per Long Ton of Sulfur									$ 57.00		$ 28.00		
Per Net Ton of Sulfuric Acid					($ 8.20)		($ 2.00)						$ 10.00
Combined Sales Value of Products				M$10,380		M$ 5,250		M$ 3,770		M$ 35,830		M$ 36,040	
All Costs Before U.S. Income Tax				6,990		6,060		7,530		34,500		27,900	
Gain or (Loss) Before U.S. Income Tax				M$ 3,390		(M$ 810)		(M$ 3,760)		M$ 1,330		M$ 8,140	

4. VALUE OF HSC FUELS AND PAYOUT

The footings of Table VIII-10 are carried forward to Table VIII-11 a 48% U.S. income tax rate is applied, the estimated remaining Gain (Or Loss) is taken to be an evaluation of the HSC Fuels according to the formula in Part VIII-1. In CASES A, D and E the values are positive and amount to a direct offset to the cost of cleaning the ROM coal.

CASE B is uneconomic, but the CASE B study is not conclusive as to the economics of intermediate concentrating technology applied to the 2% SO_2 gas of the CASE B System.

The economics of CASE C are not significant since this system is scaled to prototype dimensions.

Investment Payout is shown after interest cost in the conventional manner.

TABLE VIII-11

VALUE OF HSC FUELS, OFFSET TO CLEANING COST, PAYOUT OF INVESTMENT

	CASE A	CASE B	CASE C	CASE D	CASE E	REFERENCE
VALUE OF HSC FUELS:						
Taxable Gain (or Loss)	M$ 3, 390	(M$ 810)	(M$ 3, 760)	M$ 1, 330	M$ 8, 140	TABLE VIII-10
Gain (or Loss) After 48% Tax Effect	M$ 1, 760	(M$ 420)	(M$ 1, 960)	M$ 690	M$ 4, 230	
HSC Fuel Consumed - M Tons Per Year	640	410	300	2, 300	2, 100	FIGURE III-2
Value of HSC Fuel - Per Ton	$ 2. 75	($ 1. 02)	($ 6. 53)	$ 0. 30	$ 2. 01	
OFFSET TO COAL CLEANING:						
Clean Coal Yield - M Tons Per Year	7, 100	2, 300	920	7, 100	3, 500	FIGURE III-2
Credit (or Debit) Per Ton of Clean Coal	$ 0. 25	($ 0. 18)	($ 2. 13)	$ 0. 10	$ 1. 21	
PAYOUT OF INVESTMENT:						
Gain or (Loss) After 48% Tax Effect	M$ 1, 760	(M$ 420)	(M$ 1, 960)	M$ 690	M$ 4, 230	THIS TABLE
Depreciation	1, 634	1, 662	1, 752	7, 169	7, 214	TABLES VIII-2/6
Cash Flow	M$ 3, 394	M$ 1, 242	(M$ 208)	M$ 7, 859	M$11, 444	
Total Investment	18, 340	18, 600	29, 270	128, 520	125, 570	
Payout - Years	5. 4	15	---	16	11	

5. SHIPPING ECONOMICS

A modern steam/electric station is studiously designed and located for access to its expected fuel supply and market. Even so, the manager of a coal- or oil-fired station usually seeks to avoid dependence on a single fuel supplier, and often this can be done, especially if the fuel is oil. The design and operation of an HSC System appears to be relatively less flexible since the fuels will be expensive to ship no matter if offered by many suppliers, and a sulfur product has to be marketed in addition to energy. As to the delivery of fuel and marketing of energy no very unusual problems are foreseen. The marketing of sulfur value deserves some further comment, however:

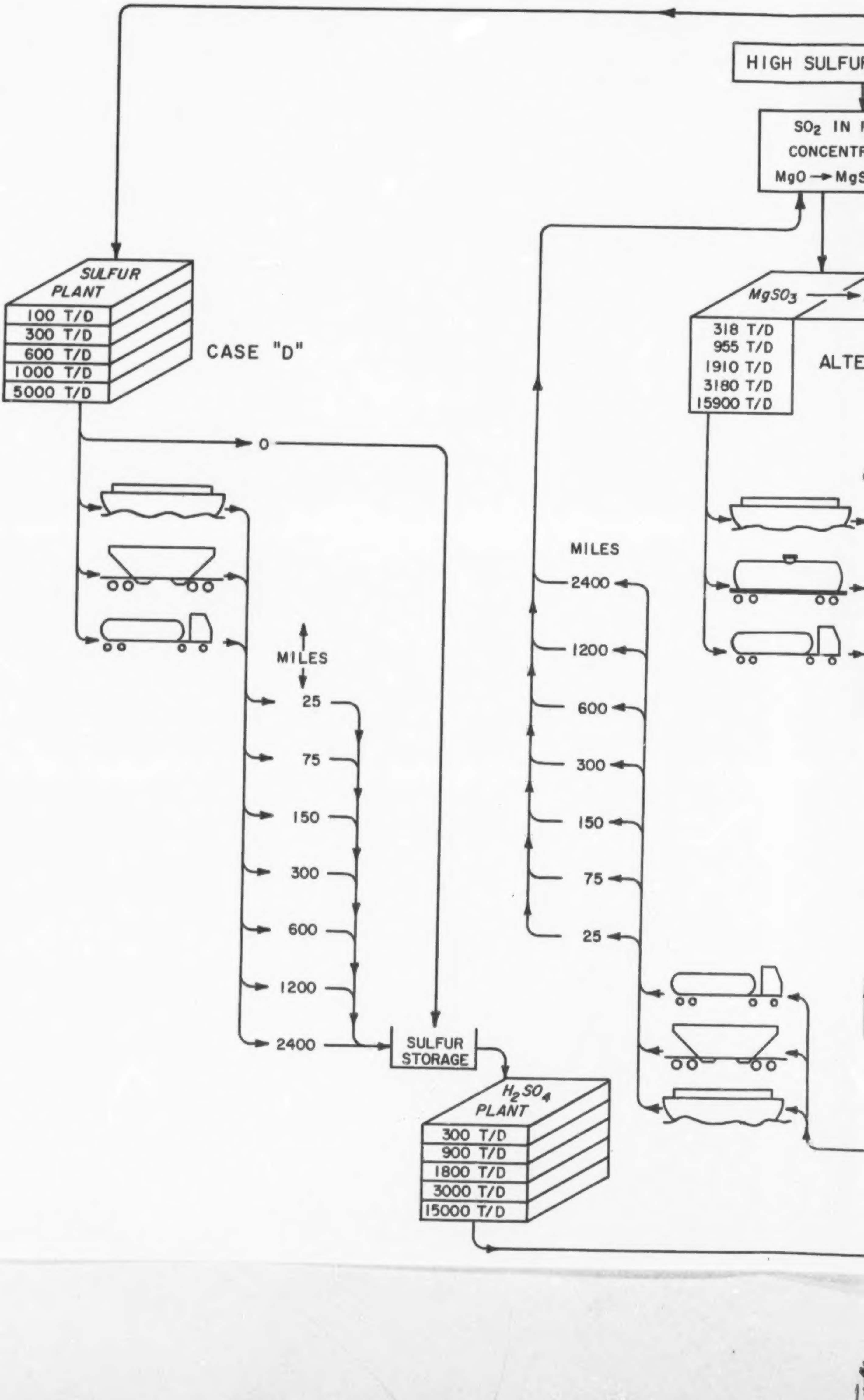
5.1 End-Product is Sulfuric Acid Delivered

In Part III-2, it was noted that sulfuric acid is the "sulfur value" of importance to this study, and since delivered price is by far the important factor in marketing sulfuric acid, the delivered cost is critical. A flue gas treating process of the type employed in CASE D offers some options for alternative processing of the dry magnesium sulfite or the 15% SO_2 gas concentrate whereby sulfuric acid can be made directly and shipped (or stored), or one of several acid intermediates can be made, shipped (or stored) and then converted to acid at the receiving end. Of these intermediates, elementary sulfur is the most important; it is cheap to ship and store, and enjoys a major status in the market. For this reason the CASE D System was presented heretofore as a sulfur making system, but since the initial process of this system can be adapted to output of any of the alternate products, it is of interest to see if "delivered cost" of sulfuric acid can be improved by traversing one or another of the alternate routes, or whether in every case the acid is best made directly from the dry magnesium sulfite.

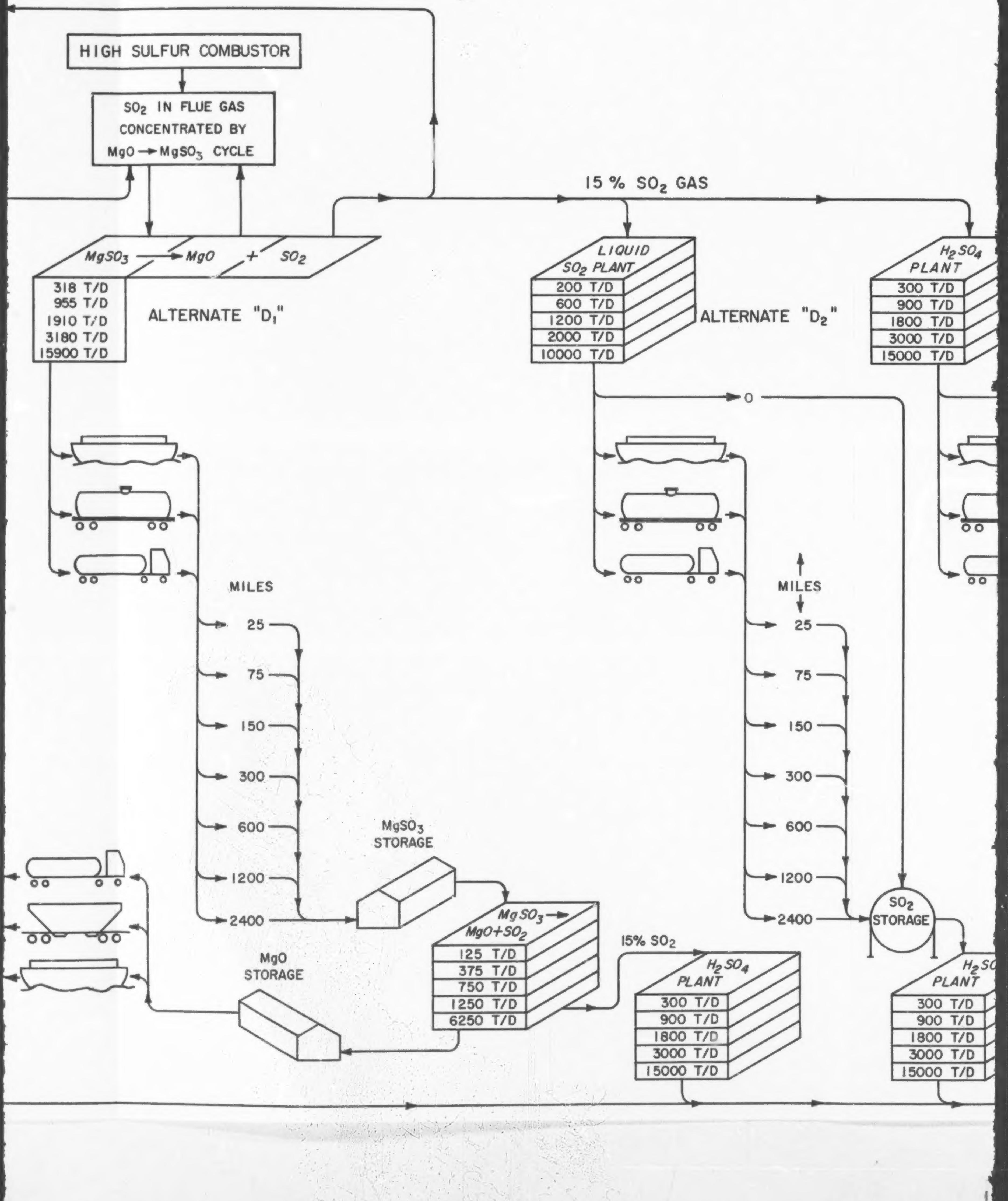
5.2 Analysis of Sulfuric Acid Delivered Cost

The question of delivered cost is generalized in the flow diagrams of Figure VIII-1. The feed material to each alternate route is dry magnesium sulfite (or more accurately "magnesium sulfur salts") which are harvested from the flue gas scrubbing cycle. Four combinations of process routes are depicted in Figure VIII-1 by which sulfuric acid can be delivered to customer's tank in each of five daily tonnages over each of seven shipping distances by each of three different modes of transport.

A


HIGH SULFUR
SO2 IN
CONCENTR
MgO → MgS
SULFUR PLANT
100 T/D
300 T/D
600 T/D
1000 T/D
5000 T/D
CASE "D"
MgSO3 →
318 T/D
955 T/D
1910 T/D
3180 T/D
15900 T/D
ALTE
0
MILES
25
75
150
300
600
1200
2400
MILES
2400
1200
600
300
150
75
25
SULFUR STORAGE
H2SO4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D

B

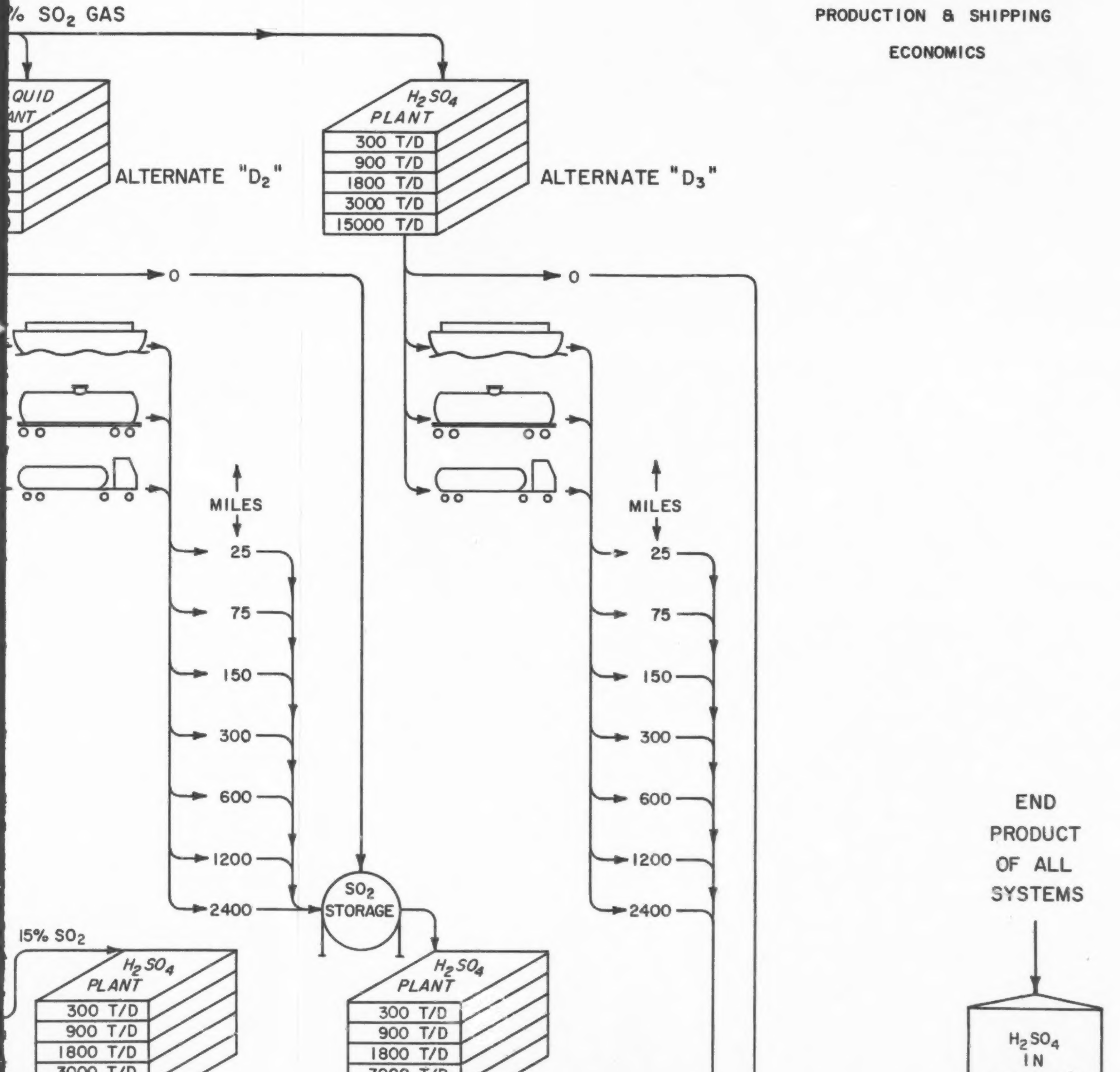

HIGH SULFUR COMBUSTOR
SO_2 IN FLUE GAS CONCENTRATED BY $MgO \rightarrow MgSO_3$ CYCLE
15 % SO_2 GAS
$MgSO_3 \rightarrow MgO + SO_2$
318 T/D
955 T/D
1910 T/D
3180 T/D
15900 T/D
ALTERNATE "D_1"
LIQUID SO_2 PLANT
200 T/D
600 T/D
1200 T/D
2000 T/D
10000 T/D
ALTERNATE "D_2"
H_2SO_4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D
0
MILES
25
75
150
300
600
1200
2400
MILES
25
75
150
300
600
1200
2400
$MgSO_3$ STORAGE
$MgSO_3 \rightarrow MgO + SO_2$
125 T/D
375 T/D
750 T/D
1250 T/D
6250 T/D
15% SO_2
MgO STORAGE
H_2SO_4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D
SO_2 STORAGE
H_2SO_4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D

C

FIGURE VIII-1

HIGH SULFUR COMBUSTOR STUDY

SULFURIC ACID & INTERMEDIATES
LINEAR MODEL FOR ANALYSIS
OF
PRODUCTION & SHIPPING
ECONOMICS


% SO_2 GAS
QUID
ANT
ALTERNATE "D_2"
H_2SO_4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D
ALTERNATE "D_3"
0
0
MILES
MILES
25
75
150
300
600
1200
2400
25
75
150
300
600
1200
2400
SO_2 STORAGE
15% SO_2
H_2SO_4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D
H_2SO_4 PLANT
300 T/D
900 T/D
1800 T/D
3000 T/D
15000 T/D
END PRODUCT OF ALL SYSTEMS
H_2SO_4 IN CUSTOMER'S STORAGE

5.2.1 Input data for production, storage and transportation costs are exhibited in Table VIII-12. Production costs include all operating and capital charges downstream of the combustor economizer in the manner of charges to sulfur product in Table VIII-5.

Storage costs are entered in Table VIII-12 to reflect barge shipping and non-barge shipping. The former costs are uniformly higher since the cost of dockage is included. The cost of transport by barge is relatively so low that this additional charge to storage is easily offset.

Transportation costs are listed in dollars per short-ton-mile for convenience, and include all elements of price as if offered by an interstate carrier. The differing rail rates for liquid SO_2 and H_2SO_4 result from the more expensive equipment required for the former.

INPUT DATA FOR PRODUCTION AND SHIPPING ECONOMICS

TABLE VIII-12

PRODUCTION COST, DOLLARS/ SHORT TON OF PRODUCT	CORRESPONDING H_2SO_4 CAPACITIES, SHORT TONS PER DAY 300.	900.	1800.	3000.	15000.
CASE D, SULFUR PLANT	108.00	80.00	73.50	68.20	61.20
H_2SO_4	41.41	29.30	26.80	24.55	21.75
D1, $MGSO_3$ AND H_2SO_4	34.78	25.05	23.35	21.82	19.90
D2, LIQ SO_2	49.30	35.80	34.00	31.50	28.30
H_2SO_4	39.73	26.80	25.20	23.22	21.00
D3, H_2SO_4	34.30	24.50	23.20	21.64	19.70

STORAGE COST, DOLLARS/SHORT TON			CORRESPONDING H_2SO_4 CAPACITIES, SHORT TONS PER DAY 300.	900.	1800.	3000.	15000.
CASE D,	SULFUR	WITH BARGE	2.64	1.13	.81	.72	.60
		W/O	1.72	.83	.66	.62	.58
D1,	$MGSO_3$	WITH	3.09	2.50	1.82	1.70	1.53
		W/O	2.69	2.25	1.75	1.65	1.52
	MGO	WITH	4.45	3.25	2.45	2.22	1.96
		W/O	3.47	2.90	2.25	2.12	1.94
D3,	SO_2	WITH	4.86	1.92	1.30	1.14	1.09
		W/O	4.29	1.72	1.21	1.08	1.07

TRANSPORTATION COST, DOLLARS/SHORT TON/MILE		DISTANCES, MILES 0.	25.	75.	125.	300.	600.	1200.	2400.
GENERAL,	BARGE	.000	.006	.006	.006	.004	.004	.003	.003
	RAIL	.000	.192	.077	.052	.035	.024	.018	.016
	TRUCK	.000	.168	.091	.079	.069	.067	.066	.051
SPECIAL,	LIQUID SO_2 USING RAIL	.000	.312	.128	.079	.053	.037	.028	.027
	H_2SO_4 USING RAIL	.000	.072	.037	.036	.032	.023	.017	.016

5.2.2 The linear program diagrammed in Figure VIII-1 is solved in the printout of Table VIII-13 in which by inspection (3rd column from the right) the minimum costs for all tonnages and distances accrue to CASE D3, e.i., direct manufacture of sulfuric acid at the site of the combustor, and shipment by barge.

RESULTS FOR HIGH SULFUR COMBUSTOR STUDY

TABLE VIII-13

PARAMETERS CAPACITIES TONS/DAY	DISTANCES MILES	UNIT COST, DOLLARS PER TON ACID CASE D,SULFUR (I=1) BARGE (J=1)	RAIL (J=2)	TRUCK (J=3)	CASE D1, MGSO3 (I=2) BARGE (J=1)	RAIL (J=2)	TRUCK (J=3)	CASE D2, LIQ. SO2 (I=3) BARGE (J=1)	RAIL (J=2)	TRUCK (J=3)	CASE D3, H2SO4 (I=4) BARGE (J=1)	RAIL (J=2)	TRUCK (J=3)
300.	0.										34.30	34.30	34.30
300.	25.	78.34	79.58	79.38	42.70	48.31	47.43	75.94	80.66	78.26	34.45	36.10	38.50
300.	75.	78.44	79.91	80.26	43.14	49.75	51.30	76.14	81.86	80.01	34.75	37.07	41.12
300.	150.	78.59	80.58	81.93	43.80	52.74	58.72	76.44	83.36	83.36	35.20	39.70	46.15
300.	300.	78.69	81.48	84.88	44.25	56.73	71.79	76.64	86.06	89.26	35.50	43.90	55.00
300.	600.	79.09	82.78	91.38	46.02	62.49	100.59	77.44	90.26	102.26	36.70	48.10	74.50
300.	1200.	79.49	85.18	104.38	47.79	73.12	158.18	78.24	97.86	128.26	37.90	54.70	113.50
300.	2400.	80.69	90.78	118.78	53.11	97.93	221.97	80.64	118.66	157.06	41.50	72.70	156.70
900.	0.										24.50	24.50	24.50
900.	25.	56.39	57.84	57.64	31.28	37.53	36.64	52.05	57.01	54.61	24.65	26.30	28.70
900.	75.	56.49	58.17	58.52	31.72	38.97	40.52	52.25	58.21	56.36	24.95	27.27	31.32
900.	150.	56.64	58.84	60.19	32.39	41.96	47.94	52.55	59.71	59.71	25.40	29.90	36.35
900.	300.	56.74	59.74	63.14	32.83	45.94	61.01	52.75	62.41	65.61	25.70	34.10	45.20
900.	600.	57.14	61.04	69.64	34.60	51.70	89.80	53.55	66.61	78.61	26.90	38.30	64.70
900.	1200.	57.54	63.44	82.64	36.37	62.34	147.39	54.35	74.21	104.61	28.10	44.90	103.70
900.	2400.	58.74	69.04	97.04	41.69	87.14	211.18	56.75	95.01	133.41	31.70	62.90	146.90
1800.	0.										23.20	23.20	23.20
1800.	25.	51.62	53.12	52.92	28.00	34.63	33.74	48.83	53.87	51.47	23.35	25.00	27.40
1800.	75.	51.72	53.44	53.79	28.44	36.07	37.62	49.03	55.07	53.22	23.65	25.97	30.02
1800.	150.	51.87	54.12	55.47	29.10	39.06	45.04	49.33	56.57	56.57	24.10	28.60	35.05
1800.	300.	51.97	55.02	58.42	29.55	43.04	58.11	49.53	59.27	62.47	24.40	32.80	43.90
1800.	600.	52.37	56.32	64.92	31.32	48.80	86.90	50.33	63.47	75.47	25.60	37.00	63.40
1800.	1200.	52.77	58.72	77.92	33.09	59.43	144.49	51.13	71.07	101.47	26.80	43.60	102.40
1800.	2400.	53.97	64.32	92.32	38.41	84.24	208.28	53.53	91.87	130.27	30.40	61.60	145.60
3000.	0.										21.64	21.64	21.64
3000.	25.	47.57	49.09	48.89	26.13	32.86	31.97	45.08	50.14	47.74	21.79	23.44	25.84
3000.	75.	47.67	49.41	49.76	26.57	34.30	35.85	45.28	51.34	49.49	22.09	24.41	28.46
3000.	150.	47.82	50.09	51.44	27.24	37.29	43.27	45.58	52.84	52.84	22.54	27.04	33.49
3000.	300.	47.92	50.99	54.39	27.68	41.27	56.34	45.78	55.54	58.74	22.84	31.24	42.34
3000.	600.	48.32	52.29	60.89	29.45	47.03	85.13	46.58	59.74	71.74	24.04	35.44	61.84
3000.	1200.	48.72	54.69	73.89	31.23	57.66	142.72	47.38	67.34	97.74	25.24	42.04	100.84
3000.	2400.	49.92	60.29	88.29	36.54	82.47	206.51	49.78	88.14	126.54	28.84	60.04	144.04
15000.	0.										19.70	19.70	19.70
15000.	25.	42.40	43.94	43.74	23.78	30.62	29.73	40.69	45.78	43.38	19.85	21.50	23.90
15000.	75.	42.50	44.27	44.62	24.22	32.06	33.61	40.89	46.98	45.13	20.15	22.47	26.52
15000.	150.	42.65	44.94	46.29	24.89	35.05	41.03	41.19	48.48	48.48	20.60	25.10	31.55
15000.	300.	42.75	45.84	49.24	25.33	39.03	54.10	41.39	51.18	54.38	20.90	29.30	40.40
15000.	600.	43.15	47.14	55.74	27.10	44.79	82.89	42.19	55.38	67.38	22.10	33.50	59.90
15000.	1200.	43.55	49.54	68.74	28.87	55.43	140.48	42.99	62.98	93.38	23.30	40.10	98.90
15000.	2400.	44.75	55.14	83.14	34.19	80.23	204.27	45.39	83.78	122.18	26.90	58.10	142.10

This result is again demonstrated in Table VIII-14, in which also the second lowest cost is shown to be:

a) Manufacture of sulfuric acid and shipment by rail on hauls up to the 300/600 mile range, and

b) Shipment of magnesium sulfur salts by barge for distances greater than 600 miles, followed by manufacture of sulfuric acid at customer's site.

TABLE VIII-14

```
                                   OPTIMAL CHOICE INDEX, WITH BARGE

I=1, CASE D,  SULFUR AND H2SO4          J=1, BARGE
  2,      D1, MGSO3 AND H2SO4             2, RAIL
  3,      D2, LIQ. SO2 AND H2SO4          3, TRUCK
  4,      D3, H2SO4

CAPACITIES DISTANCES OPTIMAL
 TONS/DAY    MILES    BEST      2ND       3RD       4TH       5TH       6TH       7TH       8TH       9TH       10TH      11TH      12TH

    300.        0.  I=4,J=1   I=4,J=2   I=4,J=3
    300.       25.  I=4,J=1   I=4,J=2   I=4,J=3   I=2,J=1   I=2,J=3   I=2,J=2   I=3,J=1   I=3,J=3   I=1,J=1   I=1,J=3   I=1,J=2   I=3,J=2
    300.       75.  I=4,J=1   I=4,J=2   I=4,J=3   I=2,J=1   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=3,J=3   I=1,J=3   I=3,J=2
    300.      150.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=3,J=3
    300.      300.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=3,J=3
    300.      600.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=4,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=3,J=2   I=1,J=3   I=2,J=3   I=3,J=3
    300.     1200.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=3,J=2   I=1,J=3   I=4,J=3   I=3,J=3   I=2,J=3
    300.     2400.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=4,J=3   I=3,J=3   I=2,J=3

    900.        0.  I=4,J=1   I=4,J=2   I=4,J=3
    900.       25.  I=4,J=1   I=4,J=2   I=4,J=3   I=2,J=1   I=2,J=3   I=2,J=2   I=3,J=1   I=3,J=3   I=1,J=1   I=3,J=2   I=1,J=3   I=1,J=2
    900.       75.  I=4,J=1   I=4,J=2   I=4,J=3   I=2,J=1   I=2,J=2   I=2,J=3   I=3,J=1   I=3,J=3   I=1,J=1   I=1,J=2   I=3,J=2   I=1,J=3
    900.      150.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=3,J=2   I=3,J=3   I=1,J=3
    900.      300.  I=4,J=1   I=2,J=1   I=4,J=2   I=4,J=3   I=2,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=2,J=3   I=3,J=2   I=1,J=3   I=3,J=3
    900.      600.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=4,J=3   I=3,J=2   I=1,J=3   I=3,J=3   I=2,J=3
    900.     1200.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=1,J=1   I=2,J=2   I=1,J=2   I=3,J=2   I=1,J=3   I=4,J=3   I=3,J=3   I=2,J=3
    900.     2400.  I=4,J=1   I=2,J=1   I=3,J=1   I=1,J=1   I=4,J=2   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=3,J=3   I=4,J=3   I=2,J=3

   1800.        0.  I=4,J=1   I=4,J=2   I=4,J=3
   1800.       25.  I=4,J=1   I=4,J=2   I=4,J=3   I=2,J=1   I=2,J=3   I=2,J=2   I=3,J=1   I=3,J=3   I=1,J=1   I=1,J=3   I=1,J=2   I=3,J=2
   1800.       75.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=3,J=3   I=1,J=2   I=1,J=3   I=3,J=2
   1800.      150.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=3,J=3
   1800.      300.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=4,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=2,J=3   I=1,J=3   I=3,J=2   I=3,J=3
   1800.      600.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=4,J=3   I=3,J=2   I=1,J=3   I=3,J=3   I=2,J=3
   1800.     1200.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=3,J=3   I=4,J=3   I=2,J=3
   1800.     2400.  I=4,J=1   I=2,J=1   I=3,J=1   I=1,J=1   I=4,J=2   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=3,J=3   I=4,J=3   I=2,J=3

   3000.        0.  I=4,J=1   I=4,J=2   I=4,J=3
   3000.       25.  I=4,J=1   I=4,J=2   I=4,J=3   I=2,J=1   I=2,J=3   I=2,J=2   I=3,J=1   I=1,J=1   I=3,J=3   I=1,J=3   I=1,J=2   I=3,J=2
   3000.       75.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=3,J=3   I=1,J=3   I=3,J=2
   3000.      150.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=3,J=3
   3000.      300.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=4,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=2,J=3   I=3,J=3
   3000.      600.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=2,J=2   I=1,J=1   I=1,J=2   I=3,J=2   I=1,J=3   I=4,J=3   I=3,J=3   I=2,J=3
   3000.     1200.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=3,J=3   I=4,J=3   I=2,J=3
   3000.     2400.  I=4,J=1   I=2,J=1   I=3,J=1   I=1,J=1   I=4,J=2   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=3,J=3   I=4,J=3   I=2,J=3

  15000.        0.  I=4,J=1   I=4,J=2   I=4,J=3
  15000.       25.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=3   I=2,J=2   I=3,J=1   I=1,J=1   I=3,J=3   I=1,J=3   I=1,J=2   I=3,J=2
  15000.       75.  I=4,J=1   I=4,J=2   I=2,J=1   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=3   I=3,J=2
  15000.      150.  I=4,J=1   I=2,J=1   I=4,J=2   I=4,J=3   I=2,J=2   I=2,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=3,J=3
  15000.      300.  I=4,J=1   I=2,J=1   I=4,J=2   I=2,J=2   I=4,J=3   I=3,J=1   I=1,J=1   I=1,J=2   I=1,J=3   I=3,J=2   I=2,J=3   I=3,J=3
  15000.      600.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=1,J=1   I=2,J=2   I=1,J=2   I=3,J=2   I=1,J=3   I=4,J=3   I=3,J=3   I=2,J=3
  15000.     1200.  I=4,J=1   I=2,J=1   I=4,J=2   I=3,J=1   I=1,J=1   I=1,J=2   I=2,J=2   I=3,J=2   I=1,J=3   I=3,J=3   I=4,J=3   I=2,J=3
  15000.     2400.  I=4,J=1   I=2,J=1   I=1,J=1   I=3,J=1   I=1,J=2   I=4,J=2   I=2,J=2   I=1,J=3   I=3,J=2   I=3,J=3   I=4,J=3   I=2,J=3
```

Since an HSC facility may well be located with no access to navigable waterway, the costs listed in Table VIII-13 were searched for minima by rail and truck haulage, with results as shown in Table VIII-15. Again sulfuric acid manufacture and shipment is indicated except for the very large tonnage and long distance at the bottom of the "Optimal Best" column, which points to sulfur as the choice for production and shipment, and conversion to acid at customer's site. Rail shipment is cheaper than truck in all instances.

TABLE VIII-15

```
                         OPTIMAL CHOICE INDEX, WITHOUT BARGE

I=1, CASE 0,  SULFUR AND H2SO4        J=1, BARGE
  2,      01, MGSO3 AND H2SO4           2, RAIL
  3,      02, LIQ. SO2 AND H2SO4        3, TRUCK
  4,      03, H2SO4

CAPACITIES DISTANCES  OPTIMAL
 TONS/DAY   MILES     BEST     2ND      3RD      4TH      5TH      6TH      7TH      8TH
    300.        0.   I=4,J=2  I=4,J=3
    300.       25.   I=4,J=2  I=4,J=3  I=2,J=3  I=2,J=2  I=3,J=3  I=1,J=3  I=1,J=2  I=3,J=2
    300.       75.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=3,J=3  I=1,J=3  I=3,J=2
    300.      150.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=3,J=3
    300.      300.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=3,J=3
    300.      600.   I=4,J=2  I=2,J=2  I=4,J=3  I=1,J=2  I=3,J=2  I=1,J=3  I=2,J=3  I=3,J=3
    300.     1200.   I=4,J=2  I=2,J=2  I=1,J=2  I=3,J=2  I=1,J=3  I=4,J=3  I=3,J=3  I=2,J=3
    300.     2400.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=4,J=3  I=3,J=3  I=2,J=3

    900.        0.   I=4,J=2  I=4,J=3
    900.       25.   I=4,J=2  I=4,J=3  I=2,J=3  I=2,J=2  I=3,J=3  I=3,J=2  I=1,J=3  I=1,J=2
    900.       75.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=3,J=3  I=1,J=2  I=3,J=2  I=1,J=3
    900.      150.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=3,J=2  I=3,J=3  I=1,J=3
    900.      300.   I=4,J=2  I=4,J=3  I=2,J=2  I=1,J=2  I=2,J=3  I=3,J=2  I=1,J=3  I=3,J=3
    900.      600.   I=4,J=2  I=2,J=2  I=1,J=2  I=4,J=3  I=3,J=2  I=1,J=3  I=3,J=3  I=2,J=3
    900.     1200.   I=4,J=2  I=2,J=2  I=1,J=2  I=3,J=2  I=1,J=3  I=4,J=3  I=3,J=3  I=2,J=3
    900.     2400.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=3,J=3  I=4,J=3  I=2,J=3

   1800.        0.   I=4,J=2  I=4,J=3
   1800.       25.   I=4,J=2  I=4,J=3  I=2,J=3  I=2,J=2  I=3,J=3  I=1,J=3  I=1,J=2  I=3,J=2
   1800.       75.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=3,J=3  I=1,J=2  I=1,J=3  I=3,J=2
   1800.      150.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=3,J=3
   1800.      300.   I=4,J=2  I=2,J=2  I=4,J=3  I=1,J=2  I=2,J=3  I=1,J=3  I=3,J=2  I=3,J=3
   1800.      600.   I=4,J=2  I=2,J=2  I=1,J=2  I=4,J=3  I=3,J=2  I=1,J=3  I=3,J=3  I=2,J=3
   1800.     1200.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=3,J=3  I=4,J=3  I=2,J=3
   1800.     2400.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=3,J=3  I=4,J=3  I=2,J=3

   3000.        0.   I=4,J=2  I=4,J=3
   3000.       25.   I=4,J=2  I=4,J=3  I=2,J=3  I=2,J=2  I=3,J=3  I=1,J=3  I=1,J=2  I=3,J=2
   3000.       75.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=3,J=3  I=1,J=3  I=3,J=2
   3000.      150.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=3,J=3
   3000.      300.   I=4,J=2  I=2,J=2  I=4,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=2,J=3  I=3,J=3
   3000.      600.   I=4,J=2  I=2,J=2  I=1,J=2  I=3,J=2  I=1,J=3  I=4,J=3  I=3,J=3  I=2,J=3
   3000.     1200.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=3,J=3  I=4,J=3  I=2,J=3
   3000.     2400.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=3,J=3  I=4,J=3  I=2,J=3

  15000.        0.   I=4,J=2  I=4,J=3
  15000.       25.   I=4,J=2  I=4,J=3  I=2,J=3  I=2,J=2  I=3,J=3  I=1,J=3  I=1,J=2  I=3,J=2
  15000.       75.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=1,J=3  I=3,J=3  I=3,J=2
  15000.      150.   I=4,J=2  I=4,J=3  I=2,J=2  I=2,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=3,J=3
  15000.      300.   I=4,J=2  I=2,J=2  I=4,J=3  I=1,J=2  I=1,J=3  I=3,J=2  I=2,J=3  I=3,J=3
  15000.      600.   I=4,J=2  I=2,J=2  I=1,J=2  I=3,J=2  I=1,J=3  I=4,J=3  I=3,J=3  I=2,J=3
  15000.     1200.   I=4,J=2  I=1,J=2  I=2,J=2  I=3,J=2  I=1,J=3  I=3,J=3  I=4,J=3  I=2,J=3
  15000.     2400.   I=1,J=2  I=4,J=2  I=2,J=2  I=1,J=3  I=3,J=2  I=3,J=3  I=4,J=3  I=2,J=3
```

6. CASE D vs. CASE D3

Since sulfuric acid is shown to be the most economical form of recovered sulfur value, it is of interest to analyze the modification of CASE D designed for output directly of sulfuric acid (CASE D3) instead of sulfur. Two SO_2 recovery trains of the general type designed for CASE A would suffice to oxidize the 15% SO_2 gas concentrate to sulfuric acid. In CASE D this concentrate is reduced to sulfur. Upstream of the gas concentrate the CASE D and D3 systems would be substantially identical, scaled to 500 MW. Estimated capital and operating cost of CASE D3 is shown in Table VIII-16. The corresponding figures for CASE D are repeated for comparison.

TABLE VIII-16

CASE D vs. CASE D3
CAPITAL COSTS, OPERATING COSTS, SALES VALUE
VALUE OF HSC FUEL

	CASE D	CASE D3
ESTIMATED CAPITAL COST		
Combustor/Boiler	M$ 15, 760	M$ 15, 760
Electric Equipment	56, 200	56, 200
Sulfur Recovery	44, 000	42, 000
Off-Sites	10, 100	10, 100
Land	1, 000	1, 000
Fixed Capital	M$128, 060	M$126, 060
Working Capital	460	460
Total Capital	M$128, 520	M$126, 520

PRODUCTION - ANNUAL BASIS		CASE D		CASE D3	
		Electricity	Sulfur	Electricity	Sulfuric Acid
Energy	MWH	4, 000, 000		4, 000, 000	
Sulfur Value	Long Ton		233, 300		
Sulfur Value	Net Ton				753, 000
SALES VALUE - UNIT		$ 7.50	$ 25.00	$ 7.50	$ 12.75
SALES VALUE - ANNUAL		M$ 30, 000	M$ 5, 830	M$ 30, 000	M$ 9, 600
OPERATING COST - ANNUAL					
Before Depreciation and Interest		6, 154	14, 055	6, 154	9, 955
Depreciation and Interest		6, 442	6, 576	6, 442	6, 423
COST BEFORE HSC FUEL CHARGE		M$ 12, 600	M$ 20, 630	M$ 12, 600	M$ 16, 400
Selling, Commercial Expense & G&A		900	409	900	670
ADJUSTMENT FOR "POLLUTION CONTROL"					
Add		14, 450		14, 450	
Deduct			14, 450		14, 450
ADJUSTED COST OF PRODUCTION		M$ 28, 000	M$ 6, 600	M$ 28, 000	M$ 2, 620
Unit Cost of Products		$ 7.00	$ 28.30	$ 7.00	$ 3.50
COMBINED SALES VALUE OF PRODUCTS		M$35, 830		M$39, 600	
ALL COSTS BEFORE U.S. INCOME TAX		34, 500		30, 570	
GAIN BEFORE U.S. INCOME TAX		M$ 1, 330		M$ 9, 000	
GAIN AFTER 48% U.S. INCOME TAX		690		4, 700	
HSC FUEL CONSUMED M TONS PER YEAR		2, 300		2, 300	
VALUE OF HSC FUEL PER TON		$ 0.30		$ 2.00	

7. RECAPITULATION

The High Sulfur Combustor, if well adapted technically to a deep cleaning operation, is evidently feasible economically provided the energy and sulfur products can be sold at, or higher than, the price levels forecast in this study. Economic netback to offset coal cleaning cost varies widely among the several "cases" analyzed. In general it is assumed that a large tonnage of sulfur processed corresponds to a large tonnage of coal cleaned. CASE A and CASE D are respectively low- and high-energy output systems that recover about the same amount of sulfur by processing the reject material from about 7 million tons per year of clean coal.

Sulfuric acid as a commercial form of recovered sulfur value appears to be much more attractive than intermediate recovery of sulfur itself, but this would be true only if the sulfuric acid moves continually to market without great price concessions. Sulfur may afford the producer some relief from a bad local market for acid in that seasonal storage and long distance shipping of sulfur can be practiced if warranted.

IX

REFERENCE LITERATURE

HIGH-SULFUR COMBUSTOR STUDY

ABSTRACTS

(A) U.S. Public Health Service. A digest of state air pollution laws, 1966. Public Health Service Publication 711, 1966, 292 pp., U.S. Govt. Printing Office, Washington, D.C. 20402, $1.50.

In all of the United States, 21 states have established distinct air pollution control comissions or boards. In other states, the functions of air pollution control are included in the duties of the Department of Health. In only two states the air quality standard has been partly specified; in other states, the measures are only in the most general terms such as laws against excessive smoke and fumes and against emissions of particles that constitute a danger to public health.

(B) Bechtel Corp. Process costs and economics of pyrite-coal utilization. A report to U.S. Dept. of Health, Education and Welfare. Contract PH 86-27-224, December, 1968, 182 pp.

The methods for recovering values from the tailings of coal beneficiation are reviewed.

For maximum desulfurization by mechanical means, it is possible: (1) to grind the run-of-mine coal to about 14-mesh, (2) to separate the ground coal into 2 fractions, (3) to treat the +28 mesh material in Baum Jigs at a separating gravity of 1.60, (4) to reject the heavy fraction as tailings, (5) to treat the minus 28 mesh material in dense-media cones and hydroclones at separating gravity of 1.45 to 1.35, (6) to reject the heavy fraction as tailings, (7) to treat the remainder by hydrocloning and froth flotation, and (8) again to reject the heavy fraction. However, operating conditions may vary appreciably as required by the nature of the particular type of coal being processed.

Mechanical methods of desulfurization remove only the pyritic S in coal. The organic S, which amounts to 30-40% of total S, would remain in the treated coal. To remove the organic S, some chemical methods, such as hydrosulfurization, must be used. The tailings from the mechanical separation methods contain a high S content in the form of pyrites. The utilization of these high S tailings is discussed. The currently known methods aim at the production of saleable SO_2, H_2SO_4, H_2S and/or elemental S. Since the tailings are not high-grade pyrites, only those processes which can handle low-grade pyrites can be used. These processes are:

(1) combustion followed by SO_2 removal, (2) carbonization followed by desulfurization, (3) solvent extraction, (4) preferential oxidation, and (5) two-stage gasification. These processes are described, and processes (1), (2) and (3) are evaluated for adoption in a commercial-scale operation. Capital requirements and operation costs are estimated. Commercial operations to utilize the S values in the high-S coal invariably involve the installation of both power generating plants and sulfuric acid plants and possibly also fertilizer plants. These are considered in detail.

(C) Paul Weir Co. (Chicago, Ill.) A feasibility study of the recovery of sulfur and iron from coal pyrites. A report to U.S Dept. of Health, Education and Welfare, Contract PH 86-65-29, May, 1966, 40 pp.; PB 170 844, Clearinghouse, Springfield, Va., 22151, $3.00.

H_2SO_4 accounts for 75% of all S consumed in the U.S. 70% of U.S. S comes from the Gulf States as elemental S from the mines. On a world-wide basis, 40% of S consumed as H_2SO_4 comes from pyritic minerals. The reserve of coal pyrites in the U.S. is estimated at about 5 billion tons in available sources. When high-S coals are processed by modern methods, the pyrite content in the coal-washing tailings is as high as 28% and averages about 15%. By fine crushing, this may reach 50%. These tailings may be further beneficiated to give a product containing S 42-48% and carbon less than 5%.

The high-S coal-washing tailings may be processed to recover S as H_2SO_4 and Fe as cintered Fe_2O_3 suitable for use by the steel makers. Processes for utilizing coal pyrites are illustrated with flow diagrams, and costs are estimated on full-scale production plants and their operations.

(D, E) Paul Weir Co. (Chicago, Ill.) An economic feasibility study of coal desulfurization. I-II. A report to U.S. Dept. of Health, Education and Welfare, Contract PH 86-65-29, Oct., 1965, 57 + 130 pp.; PB 176 845, 176 846, Clearinghouse, Springfield, Va., 22151. $6.00.

Table 1

Coal consumption and use pattern in the U.S. 1,000 tons

	1960	1962	1964*
Electric utilities	173, 882	190, 833	226, 000
Coke and gas plants	81, 015	74, 262	89, 000
Retail dealers	30, 405	28, 188	20, 000
Others	95, 127	94, 491	100, 000
Total (bituminous and lignite)	380, 429	387, 774	435, 000
Anthracite	17, 600		14, 000

* Preliminary

The sulfur content of the various sources of coal in the U.S. is presented in detailed localities and coal seams. They range from 0.3 to 7.6%, but most commonly between 0.8 and 4.5%. The high sulfur coals, e.g., containing 2.0 to 7.6% sulfur, were further examined for the form of sulfur, i.e., pyritic, organic and sulfate. In general, the sulfate form is very small (0.01-0.13%), but pyritic sulfur accounts for about 2/3 of the total, with the remainder in the form of organic sulfur.

Crushing the coal followed by gravity separation (e.g., washing with water) removes a large part of pyritic sulfur but not the organic sulfur. For example, a West Virginia coal containing total sulfur 4.05% (pyritic 2.59, organic 1.44 and sulfate 0.02%) is crushed to minus 30-mesh, and washed to discard 22.6% tailings, gave a product (77.4%) containing 2.31% total sulfur. Thus, the reduction of sulfur content was about 43%. The sulfur removal efficiency decreases rapidly as the size of the crushed coal increases. For example, the same West Virginia coal when crushed to minus 1-1/2 inch pieces gave a yield of 82.8%, a sulfur content of 3.9% in the product (a reduction of S of only 3.7%).

About one-half of all coals consumed in the United States is burned for power generation. In the majority of cases, no attempt was made to reduce the sulfur content. Coal washing, however, is generally applied to coals destined for metallurgical coke making or for gas production. The cost of coal washing including cost of crushing to minus 14-mesh is estimated at about $0.42 per ton of product (based on a yield of about 79%). The efficiency of sulfur removal by washing varies greatly with different sources of coal on account of different nature of sulfur in the coal.

Case studies are given for the partial removal of sulfur from several coals containing more than 2% total sulfur, including coals from Illinois, West Kentucky, Ohio and West Virginia. Process flow diagrams are given in each case.

(F) Arthur D. Little, Inc. (Cambridge, Mass.) A study of process costs and economics of pyrite-coal utilization. A report to U.S. Dept. of Health, Education and Welfare, Contract PH 86-27-258, Mar., 1968, 266 pp.; PB 182 303, Clearinghouse, Springfield, Va., 22151, $3.00.

In order to meet the requirement of air pollution control in densely populated localities in the U.S., it would be more economical to remove the S in the fuel at the source than to remove the SO_2 in the flue gas after burning the high-S fuel. In the case of high-S coal, an economic gain is possible by treating the coal to decrease its S content to below 1% and converting the pyrites in the tailing to H_2SO_4. Detailed estimates were made on the costs of S removal from coal, H_2SO_4 manufacture from the pyritic refuse from coal, and iron oxide recovery from the pyrite cinders. Full utilization of the pyritic refuse includes a large-capacity phosphate fertilizer plant to absorb the H_2SO_4 produced. Suitable plant locations for the whole project and the capital investments are discussed.

(G) Spencer, J.D. Review of Bureau of Mines coal program, 1968. U.S. Bur. Mines, Inf. Circ. 8416, 1969, 94 pp.

In the field of coal research the U.S. Bureau of Mines in 1968 was continuing in the following fields: (1) Coal mines environment studies; (2) coal mining technology; (3) coal mine hazards and safety; (4) coal cleaning; (5) coal transportation; (6) coal burning methods; (7) coking and carbonization; (8) coal gasification; (9) chemical products from coal; (10) chemistry and structure of coal; (11) Analytical and testing methods for coal.

(H) U.S. Office of Coal Research, Annual Report 1969. U.S. Dept. Interior, Office of Coal Research, 1969, 65 pp.; Govt. Printing Office, Washington, D.C. 20402, $0.70.

Activities of the Office of Coal Research during the calendar year 1968 are briefly reported covering coal mining, coal preparation, coal and lignite gasification, ash utilization, and sulfur dioxide control in furnaces.

(J) Mining Information Services. U.S. coal production by company, 1968. McGraw-Hill, Inc., New York, N.Y. 1969, $5.00.

Major coal-producing companies in the U.S. who during 1968 produced 100,000 tons or over, are listed with the location of mines and production figures. There are about 630 such companies. U.S. production of anthracite in 1968 was 11,025,482 tons and bituminous and lignite together 545,000,000 tons.

(K) National Coal Assoc. Steamelectric plant factors, 1968, 8th ed., 110 pp., National Coal Association, Washington, D.C. 20036, $5.00.

The statistics of U.S. power plants burning coal, oil, and natural gas are presented. Operating data of 397 plants are tabulated with information on capacity, "dependable power", net generation (kwh for the year), fuel consumed, cost of fuel, ($/MM Btu). Also given is a list, by city and state, of steamelectric power plants under construction and in the planning stage.

(L) EBASCO Services, Inc. 1968 Business and economic charts. Ebasco Industries Co., New York, N.Y., 1969, 45 pp.

Statistics are presented in graphic form covering the following subjects:

1. U.S. Population.
2. Gross national product
3. New home building
4. Industrial capital outlays
5. Power plants
    a. Public
    b. Private
    c. Cooperative

6. Power plants capacities, 1948-1969
7. Electric power consumption
8. Energy sources: Coal, oil, gas, hydro, nuclear
9. Utility companies financial sources
10. Utility company taxes
11. Power generating costs
12. Power transmission costs
13. Power rates

(M) Bishop, J.W., Robinson, E.B., Ehrlich, S., Jain, A.K., and Chen, P.M. Status of the direct contact heat transferring fluidized bed boiler. Paper presented at the ASME Winter Meeting, New York, N.Y., December 1-5, 1968, Paper No. 68-WA/FU-4, 13 pp.; Am. Soc. Mech. Engrs., New York, N.Y. 10017, $1.50.

The recent fluidized bed boiler development work sponsored by the U.S. Office of Coal Research involves replacement of the conventional boiler furnace with fluidized suspension of inert material (e.g. sintered coal and crushed and screened to a uniform suitable particle size) into which coal is injected and burned. The water tubes and superheaters are suspended at suitable levels in the fluidized bed. High heat releases and heat transfer direct from bed material to heating surface obtained by this method result in very high steaming capacities from an exceptionally small boiler. Because of the extremely high heat transfer rates, an upper limit must be set for the heat exchange surface to oxidizing fluidized bed volume ratio, otherwise the flame would be quenched and operation would be unstable. From experimental data derived in operation of a full-scale single-module boiler, packaged railroad transportable units can be built up to 300,000 lb/hr capacity or larger. The envisioned utility boilers of 2,000,000 lb/hr and larger, represent about 15% of the overall size of a similar-capacity pulverized coal unit. Envisioned large cost savings should make coal more competitive as a boiler fuel. The use of limestone for sulfur-oxide abatement in this system is far more effective than the conventional boilers. Reductions in about 65% of SO_2 emissions have already been accomplished and greater reductions are anticipated.

(N) Walker, F.E., and Hartner, F.E. Forms of sulfur in U.S. coals. U.S. Bureau of Mines, Inf. Circ. 8301, 1966, 51 pp., $0.35.

The sulfur contents of approximately 2,900 samples of coals from 30 states in the U.S. have been analyzed and are reported as organic, pyritic, and sulfate sulfur, respectively. The coals are classified under 11 types such as anthracite, semi-anthracite, bituminous, subbituminous, lignite, etc.

(P) Chemical Construction Corp., (New York, N.Y.). Report to Bituminous Coal Research, Inc., Investigation of processes to remove SO_2 and recover sulfur compounds as salable products from boiler flue gases. June 20, 1968.

Exploratory investigation of SO_2 control processes, preliminary design and economic evaluation, intended to highlight relatively promising areas for further study.

(Q) MSA Research Corp. (Evans City, Pa.) and Singmaster and Breyer (New York, N.Y.). Inorganic liquids for removing SO_2 from flue gases. Phase I. National Air Pollution Control Administration, Interim report, April 9, 1969, in connection with contract PH 22-68-11, 181 pp.; PB-183974, Clearinghouse, Springfield, Va. 22151, $3.00.

A literature survey was made on the absorption of SO_2 from fuel gas by inorganic liquids in the temperature range 200-600° F, covering molten salts and molten alloys. Experimental work was carried out on the following proposed methods for SO_2 removal: (1) A Pb-Sn eutectic was used as the absorbent at 400° F. The SO_2 produces a dross of sulfates. The dross was separated and heated at a higher temperature to decompose the sulfate and to regenerate the Pb-Sn alloy. (2) Limestone powder was mixed with a molten eutectic KNO_3-$LiNO_3$ and was used as SO_2 absorbent at 250° F. $CaSO_4$ was produced in the process. It was filtered off and discarded. (3) CaO was dissolved in molten KNO_3-$LiNO_3$ eutectic, and the molten solution was used as the absorbent of SO_2. The pregnant absorbent is heated to decompose $CaSO_3$ and to regenerate the CaO in the molten solution. (4) $MgSO_4$ was dispersed in a molten entectic system and was used as the absorbent for SO_2 and SO_3. It was regenerated by steam stripping. Preliminary economic evaluation was made for each of the four proposed methods based on a hypothetic power plant of 800 megawatt capacity. Operating costs thus estimated seem to be in the range of feasibility for some of the proposed methods.

(R) Arthur G. McKee & Co. (San Francisco, Calif.) Systems study for control of emissions of the primary nonferrous smelting industry. National Air Pollution Control Administration, McKee Report No. 993; Final report under contract PH-86-65-85; June, 1969. In three vol., 187 + 180 + 218 pp. resp.; PB-184 884-5-6, Clearinghouse, Springfield, Va., 22151, $9.00.

Some 2, 800, 000 tons of sulfur are contained in the oxide gases generated annually at the smelters in the U.S. About 31% of this sulfur is recovered mostly as sulfuric acid. The remaining 1, 920, 000 tons per year are emitted to the atmosphere. Copper smelters are the source of 76.5% of this. Lead and zinc smelters emit the balance. Over 97% of all emissions are from smelters west of the Mississippi River. The problems of reducing the emissions from the nonferrous smelters seem at present very difficult economically.

Neither now nor in the period up to at least 1975 can all of the potentially recoverable sulfur in the west be sold as sulfuric acid. All of it might be sold if a portion could be converted from sulfur oxides to elemental sulfur at a cost that is low enough to be competitive. The area of Arizona-New Mexico-West Texas has the largest potential production of sulfur by-products but a relatively small market for them, especially for sulfuric acid.

Production of sulfuric acid from the more concentrated gases can be profitable where a market for acid exists. The production cost for converting sulfur oxides in the gases to elemental sulfur by modern processes is not yet defined. Production of sulfur for sale at competitive prices will probably be possible only by recovery from gases containing high concentrations of sulfur oxides.

Therefore, the major technical and economic problem is the treatment of offgases containing low concentrations of sulfur oxides. Present technology is not adequate to economically recover sulfur oxides from these weak offgases. Economic emission control seems to lie in the direction of process or practice changes that have the effect of curtailing generation of weak offgases and delivering other offgases to sulfur oxide conversion units at the highest feasible concentration.

(S) Federal Power Commission. Air pollution and the regulated electric power and natural gas industries. Federal Power Commission Staff Report, September 1968, Washington, D.C., 366 pp.

The overall picture of energy sources, supplies and consumption in the U.S. is presented for the period 1937-1965 and projected to 1980. The urgency of environmental care and management calls for surveillance of the quality of fuels, such as coal, coke, and oil, which are burned to produce heat and electricity. Toward this objective legislation is now proceeding at federal, state and municipal levels to set standards for air quality and to control the amounts of air pollutants, especially sulfur oxides both in the fuels delivered to the locality and in the gases discharged from burners and furnaces, and in the air in the vicinity of fuel burning installations. Cities that already have laws for air pollution control are listed. In order to comply with these laws, the fuel-burning and waste material-burning establishments, such as power stations, apartment houses, and incinerator plants, must either install sulfur oxides (and nitrogen oxides, hydrochloric acid etc.) removal devices, or limit the sulfur content of the fuel used. Just what is to be done to improve the existing air quality and to maintain a certain standard, and the costs involved are discussed. A study of a specific case, the St. Louis, Mo., area has been made and the results are presented in detail.

(T) Helfinstine, R.J., Shimp, N.F., and Simon, J.A. (Illinois State Geological Survey, Urbana, Ill.) Sulfur varieties in Illinois coals - Float-sink tests. U.S. Public Health Service Contract No. PH-86-67-206, Report August 10, 1969; 9 + 87 pp.

Results of laboratory tests and analyses of some 1,200 samples of coal from more than 800 mines in the State of Illinois are reported in 8 tables and 240 graphs. Special attention was paid to the chemical and physical properties, sulfur content, (sulfate, organic and pyritic), washability and ash fusion temperature. The run-of-mine coal contains total sulfur from below 1.5% to as high as 6.99%.

Each sample was divided into float and sink fractions as in coal washing. Analyses of the float and sink fractions indicated that only about 1/4 of the Illinois coal can be washed to give a product at 80% recovery containing 2.5% S or less. Only a few mines give coal that can be washed to give a practical percentage of recovery and a S content of 1.5% or less.

The float fraction usually had less sulfur in the smaller particle sizes, but the differences were not great enough to make fine grinding a practical means of sulfur reduction.

The maximum sulfur in the 1.60 specific gravity sink fraction for the 40 Illinois coal samples tested was 26%. Five of the 40 samples testd had more than 20% sulfur in the 1.60 sink fraction. These sink products indicate that the mine refuse from some Illinois mines is a potential source of sulfur.

(U) National Air Pollution Control Administration. Air quality criteria for particulate matter. U.S. Dept. of Health, Education and Welfare, National Air Pollution Control Administration, Publication No. AP-49, January 1969, Washington, D.C., 211 pp.

By the Congressional Air Quality Act of 1967 the Secretary of Health, Education and Welfare has the responsibility of issuing to the States criteria of air quality to guide them in the legislation and promulgation of laws for the protection of people from adverse effects of air pollution.

This report is concerned chiefly with particulate matter in ambient atmosphere, its sources, its effects on health, both human and animal, on vegetation, and on materials and buildings, and the desirability of its abatement and control for health, economic and aesthetic reasons. It is observed that under the conditions prevailing in localities where studies were made, adverse effects were noted when annual geometric mean levels of particulate matter exceeded 80 micro g./cu.m. for health, and 60 micro g./cu.m. for materials and buildings. Visibility reduction to about 5 miles was observed at annual mean exceeding 150 micro g./cu.m. It is recommended that when setting up ambient air quality standards, consideration should be given to requirements for margins of safety which take into account long-term effects on health and materials occurring below the above mentioned levels. With subject index.

(V) National Air Pollution Control Administration. Air quality criteria for sulfur oxides. U.S. Dept. of Health, Education and Welfare, National Air Pollution Control Administration Publication No. AP-50, January, 1969, Washington, D.C., 178 pp.

As a part of the Air Quality Criteria issued to the States by the Department of Health, Education and Welfare in accordance with the Air Quality Act of 1967, this report concerns chiefly with sulfur dioxide in ambient atmosphere, its sources, its effects on health, both human and animal, on vegetation, and on materials and buildings, and the desirability of its abatement and control for the protection of people and their health and for environmental wholesomeness. Furthermore, in addition to health considerations, the sulfur dioxide in the air must be abated and controlled for economic and aesthetic reasons. It is noted that under the conditions prevailing in localities where studies were made, adverse health effects were observed when 24-hour average levels of SO_2 exceeded 300 micro g. /cu. m. (0.11 ppm). Visibility reduction to about 5 miles was observed at 285 micro g. /cu. m. (0.10 ppm); adverse effects on materials and buildings were observed at an annual mean of 345 micro g. /cu. m. (0.12 ppm); and adverse effects on vegetation were observed at an annual mean of 85 micro g. /cu. m. (0.03 ppm). It is recommended that when setting up ambient air quality standards, consideration should be given to requirements for margins of safety which take into account long-term effects on health, and materials occurring below the above-mentioned levels. With subject index.

(W) National Air Pollution Control Administration. Control techniques for particulate air pollutants. U.S. Dept. of Health, Education and Welfare, National Air Pollution Control Administration, Publication No. AP-51, January 1969, 215 pp.

In accordance with the Congressional Air Quality Act of 1967, the U.S. Department of Health, Education and Welfare has compiled this manual of control techniques for particulate air pollutants. It encompasses the control techniques at the sources of emission, methods and equipment for removal of particulate matter from air discharge streams, disposal of collected particulate wastes, capital costs of equipment and operating costs. Among the various types of particles removing equipment described are baffles, cyclones, entrainment removers, mist eliminators, high and low-voltage electrostatic precipitators, fabric filters and afterburners. With subject index.

(X) National Air Pollution Control Administration. Control techniques for sulfur oxide air pollutants. U.S. Dept. of Health, Education and Welfare, National Air Pollution Control Administration Publication No. AP-52, January 1969, 122 pp.

In accordance with the Air Quality Act of 1967, the U.S. Dept. of Health, Education and Welfare has compiled this manual of control techniques applicable to the sulfur oxides in the atmosphere. It encompasses the control techniques at the source, processes, methods, and equipment for removal of sulfur oxides from exhaust gases, capital costs of equipment and operating costs. Among the various processes described are the alkalized alumina process, the limestone and dolomite injection process, the Beckwell SO_2 recovery process, and the catalytic oxidation process. The evaluation of these processes is discussed. With subject index.

(Y) Babcock & Wilcox Co. Air pollution control program. Babcock & Wilcox Brochure BR-905, June 1969, 8 + 2 pp.

A brochure addressed to utility companies and other interested groups soliciting financial support for a comprehensive test program of a system of air pollution control for power boilers. A preliminary study comparing several proposed processes has convinced Babcock & Wilcox Co. that the process proposed by Esso Research & Eng. Co. using a dry regeneratable sorbent would fit into a system (designed as B&W-Esso system) that would fulfill a long-term need of the utility companies. Babcock & Wilcox is proposing a four-year development program in three stages or phases. Phase I: Process designing, sorbent material testing, sorbent regeneration technique improvement. Phase II: Pilot plant erection (25 Mw capacity), pilot plant operation, and sulfur oxides recovery study. Phase III: Prototype plant designing based on an 800 Mw coal-fired power station, erection of the prototype plant, and operating it.

The goal of Phase III is to achieve a particulate pollutant removal of 99%, a sulfur oxides removal of 90%, a nitrogen oxides reduction, the production of byproduct H_2SO_4 and a low overall cost of the system (capital and operating). An estimate of the financial requirements ($6, 750, 000) for the program is given.

(Z) Nat'l Air Pollution Control Admin. A status report: process control engineering, R&D for air pollution control. U.S. Dept. Health, Education and Welfare. Nat'l Air Pollution Control Admin., Cincinnati, O., Nov. 1969, 45 pp.

The Nat'l Air Pollution Control Admin. through its Process Control Engineering Section (PCE) continued in 1969 its programs of research and development in the field of controlling emissions of SO_2 and other pollutants from stationary sources. Specific studies have been undertaken in collaboration with other organizations; the principal ones are the following:

1. Limestone injection to control SO_2: with Tennessee Valley Authority.
2. Removal and utilization of pyrite sulfur in coal: with U.S. Bur. Mines.
3. Fluidized bed combustion combined with limestone additive: with Office of Coal Research, and U.S. Bur. Mines.
4. Meteorological research related to air pollution: with Environmental Science Services Admin.
5. Power industry needs: with Atomic Energy Commission.
6. Using "red mud" as absorbent for SO_2 in flue gas: with the City of Chicago.
7. Coal reserves and sulfur in the coals: with the States of Illinois, Ohio, Montana, New Mexico, Utah, Indiana and Mississippi.
8. Washability of Illinois coal: with the State of Illinois.

As of June 30, 1969, PCE had 47 active contracts with private firms and institutions outside of the Federal government to carry out specific studies and research.

Internationally, PCE has an agreement with the National Coal Board of England for exchange of basic information on fluidized-bed combustion of coal.

(AA) National Air Pollution Control Admin. The cost of clean air. First report of the Secretary of Health, Education and Welfare to the Congress, June 1969, 111 pp.; U.S. Dept. Health, Education and Welfare, Washington, D.C., NAPCA-FS-1.35:969.

For the fiscal years 1970-1974, the costs of the air pollution control program in the United States, covering government expenditures at the federal, state and local levels, are estimated as follows:

	Millions of Dollars
1969	122.9
1970	159.6
1971	207.4
1972	269.3
1973	350.0
1974	454.5
	1,563.7

These expenditures comprise research and development as well as abatement and control.

For expenditure on the part of industry based on substituting low-sulfur coal for high-sulfur coal and substituting 1%-sulfur fuel oil for high-sulfur fuel oil, along with maximum control of particulate emissions, the total cost of air pollution control in 85 metropolitan areas are estimated as follows, the low and high estimates indicating differences in costs of equipment and in operating costs involved in the fuel desulfurization processes and in flue gas treatment processes:

	Fossil fuel burning power plants		Chemical, metallurgical and manufacturing plants	
	(Millions of dollars)			
	Low	High	Low	High
1971	401.6	455.0	41.7	82.3
1972	635.2	730.8	73.3	137.4
1973	662.1	766.7	74.5	139.8
1974	689.3	801.4	76.5	140.8

It is noted that the highest estimate of annual costs for controlling sulfur oxides and particulate emissions from central power stations is less than 0.5% of the operating revenues, and that of industrial plants less than 2% of the value of shipments from such plants. The only exception is the sulfuric acid industry, in which case the "high" estimate for the year 1974 will be 12.66% of the value of shipments excluding credit for values recovered in connection with emission control measures. This credit could substantially off-set the cost of emission control.

(BB) Rallo, T.V. (Foster Wheeler Corp., Livingston, N.J.) Technology and economics for domestic boiler and power plant designs. Foster Wheeler Corp. Report, June 12, 1969; Contract 15 + 24 pp.

The technology and economics of steam power plants in size range of 50-400 Mw burning normal fuels have been studied with a view of determining such designs as most amenable to combination firing of refuse and fossil fuels. Design concepts of ten examples in the said size range are presented illustrating the state-of-the-art in the United States. The boilers of the ten power plants are: 2 oil-, 3 gas-, and 5 coal-fired; steam rates range 500-2, 390 M lb. /h.; superheat steam temperature 950-1, 050° F, pressure 1, 275-2, 486 psig; boiler efficiency 84. 23-89. 32%; total plant costs including turbine-generator equipment accessories and structures $203-108 per kw, resp., electricity production costs $0. 01043-0. 00645 kwh., resp., including operating costs, fuel costs, and fixed charges. Detailed analyses of these figures are given. Sketches of the ten steam generators are shown. A special burner, the FW multi-fuel burner, is also shown. This burner is adaptable to all kinds of fuels including gas, oil, and coal, burned interchangeably or simultaneously.

(CC) Zimmerman, R.E. (Paul Weir Co., Chicago, Ill.) The economics of desulfurization coal by conventional coal preparation methods. Paper No. 24D presented at the 59th National Meeting of A.I.Ch.E., Columbus, O., May 15-18, 1966, 18 _ 9 pp.

Various methods in conventional coal washing are investigated. The washing operation reduces the sulfur content of coal by removing a part of the pyrites and marcasites which constitute the pyritic sulfur, as against organic sulfur which is immune to washing operation. But the organic sulfur may amount to as much as 50% of total sulfur in coal. Therefore, sulfur removal by washing has substantial limitations. Within the limitations, the effectiveness of washing improves as the size of the coal particles decreases and as the specific gravity of the separation is lowered. But this improvement is only possible at the expense of coal recovery. If low-sulfur coal is sold at a premium, that premium is balanced by the cost of the washing operation plus the cost of combustible values left in the coal refuse, that is, the lower the sulfur content the greater the amount of coal refuse. Capital and operating costs of coal washing are illustrated.

(DD) Deurbrouck, A.W., and Palowitch, E.R. Survey of sulfur reduction in Appalachian region coals by stage crushing. U.S. Bur. of Mines Inf. Circ. 8282, 1966, 37 pp.

The sulfur, particularly the pyritic form, in coal can be reduced to a certain extent by crushing the coal to liberate the pyrite-rich particles followed by gravity separation in aqueous medium. This method was applied to the coal from the Appalachian region in Pennsylvania, Ohio and Tennessee. Lab. tests show that, when the coal is crushed to pass 14-mesh, followed by washing, the sulfur content of the product is significantly lower than in the original coal (e.g. 1.5% vs. 3.0%) while coal recovery is about 84%. However, there are many exceptions to these figures, depending on the location of the coal beds, even if they are in the same general region.

(EE) Dennis, R., and Bernstein, R.H. (resp., GCA Corp., Bedford, Mass., and Chas. T. Main, Inc., Boston, Mass.) Engineering study of removal of sulfur oxides from stack gases. - Final report. Am. Petr. Inst., Air and Water Conservation Committee Report, Aug. 1968, 69 + 5 pp.

Four processes for the removal of SO_2 from the powerplant stack gases were selected for detailed study from the economic and engineering standpoint. They are: (1) Dry dolomite (limestone) injection followed by wet scrubbing, (2) Catalytic oxidation followed by absorption in water, (3) Alkalized alumina process, and (4) Reinluft process. As a result of this study, the capital investment and operating costs of SO_2 removal are estimated based on an oil-fired power plant rated at 800 Mw and operated at 60% load factor, as follows:

	PROCESS			
Cost	Dolomite Injection-Wet Scrubbing	Catalytic Oxidation	Alkalized Alumina	Reinluft
Capital Cost				
10^6 $	0.51	14.0	10.0	13.1
$/kW	0.50-0.63	17.50-19.00	11.50-13.00	16.50-17.00
Operating Cost				
Mills/kWhr	0.20	0.59	0.73	0.58
$/bbl oil	0.14	0.41	0.51	0.41

(FF) Argenbright, L.P., and Preble, B. (Arthur G. McKee & Co., San Francisco, Calif.) Sulfur oxides from western smelters. I. Quantities, sources, and costs of recovery. Paper presented at the 158th Nat'l Meeting of Am. Chem. Soc., New York City, Sept. 7-12, 1969, 17 + 9 pp. (Based on a study performed for National Air Pollution Control Administration under Contract PH-86-65-85.)

Some 2.2 million long tons per year of sulfur is contained in the sulfur oxide gases generated in the operation of copper, zinc, and lead smelters in the western U.S. Nearly 23% of this is recovered, almost all as slufuric acid. More acid can be made where markets are available, but many smelter gas flows are too dilute for economic acid manufacture. Costs for concentration of these gases by presently known processes are excessive. Estimated costs of sulfuric acid made from the smelter gases are given based on the SO_2 concentration in these gases. These estimated costs are compared with that of sulfuric acid made from elemental sulfur. Development of improved smelting processes that would produce more concentrated gases could eventually lead to recovery of over a million tons of sulfur values now wasted. Costs are given for sulfur oxide recovery by the Cominco process and by the Asarco process, respectively. The Cominco process ends up with concentrated SO_2, and the Asarco process ends up with liquid elemental sulfur. It seems a low cost process is needed to convert the SO_2 from the smelters to elemental sulfur, which could be more economically stored and shipped than equivalent quantities of sulfuric acid.

(GG) Yodis, A.W., Boucher, S.B., Edgecomb, R.H., Falk, G.B., Harrer, T.S., and Park, R.S. (Allied Chem. Corp., New York, N.Y.) Applicability of reduction to sulfur techniques to the development of new processes for removing SO_2 from the flue gases. National Air Pollution Control Administration, Contract No. PH 22-68-24, Interim Report Sept. 26, 1969; 379 - 4 pp.

The literature related to the reduction of SO_2 to elemental sulfur is reviewed. The established processes were studied with the object of economic evaluation if they should be selected for the treatment of flue gases containing SO_2. These studies cover the following processes:

(1) Hot carbon reduction
(2) Reduction with methane
    (a) The Asarco process
    (b) The West process
    (c) Kulcsar process
(3) Reduction with $CO + H_2$ mixture
(4) The normal Claus process
(5) The low-temperature Claus process

Thermodynamic and equilibrium calculations for the above processes are presented. Industrial plant-scale layouts are proposed. Capital investment and operating costs are estimated.

(HH) Aerojet-General Corporation. Applicability of aqueous solutions to the removal of SO_2 from flue gases, National Air Pollution Control Administration, Contract PH 86-68-77, Final Report Volume I, October 1970.

The feasibility of using aqueous systems for removing SO_2 from flue gases is appraised. Assessment of methods, literature survey, preliminary economic evaluation, selection of candidate processes are included. Thirty processes are identified and 22 evaluated. The economic evaluation basis is a flue gas containing 0.3 vol. % SO_2 from a coal burning steam-electric station of 120 MW capacity.

(JJ) a A Bituminous Coal Research, Inc. Report. An evaluation of coal cleaning processes and techniques for removing pyritic sulfur from fine coal. National Air Pollution Control Administration Report 1970 (BCR-L-339; Contract No. PH-86-67-139), 265 + 17 pp.

The methods of pyritic sulfur removal from coals were investigated based on 70 utility coals from various seams in Illinois, eastern Ohio and western Pennsylvania. These methods include (1) compound water cyclone, (2) concentrating spiral, and (3) concentrating table. Tests were made on two fine sizes of coal (a) minus 30 mesh (corresponding to recycle of a utility pulverizer) and (b) p.c. grade (corresponding to the "as fired" pulverized coal). The results of the tests show that the pyritic sulfur in No. 6 Ohio coal ground to minus 30 mesh can be reduced 93.7% and for Illinois No. 6 coal the reduction was only 51.2% at the same particle size range. At the p.c. grind, the Pennsylvania Lower Freeport coal showed a pyritic sulfur reduction of 95.0%, but the Pennsylvania Middle Kittanning coal only 54.5%. The pyritic sulfur in the coals tested vary greatly even from the same seam; for example, the Middle Kittanning seam showed a low of 0.22% at Clarion County and a high of 8.45% at Clearfield County, r.o.m., minus 1.5 inch basis. The adaptability of the 3 methods of coal cleaning was found to vary with the particular types or sources of coal. The method of air classification was briefly tested; results are preliminary.

(JJ) b Bituminous Coal Research, Inc., Final Report. An evaluation of coal cleaning processes and techniques for removing pyritic sulfur from fine coal. National Air Pollution Control Administration. Contract No. PH-86-67-139.

Material contained in the earlier report, (JJ)a see abstract above, is repeated in somewhat abbreviated form together with data on washability of 20 additional coals from "Southern, Western and Mid-Western" states. But the majority of the 20 additional coals "were quite different from the 70 coals" earlier evaluated. Of the 20 coals, the few that were high in pyritic sulfur were high also in organic sulfur "so that total sulfur reductions obtained were not as impressive as previously reported reductions".

(KK) a Saltsman, R.D. (Bituminous Coal Research, Inc., Monroeville, Pa.) The removal of pyrite from coal. Paper presented at Am. Soc. Mech. Engrs. Annual Meeting, New York, N.Y., Dec. 1-5, 1968; ASME preprint 68-WA/FU-2, 8 pp.

The occurrence of pyrite in coal is discussed and a brief review is made of previous work at Bituminous Coal Research, Inc., on pyrite removal that led to two substantial projects. The first of these projects, cosponsored with 12 Eastern utilities, is a program to evaluate methods for optimizing pyrite removal from the pulverizing mill prior to combustion. The second project is sponsored by the U.S. Public Health Service and is an evaluation of coal cleaning methods and techniques for removal of pyritic sulfur from fine coal. Results from these projects are discussed.

(KK) b Saltsman, R.D. (Bituminous Coal Research, Inc., Monroeville, Pa.) The removal of pyrite from selected Pennsylvania coals. Paper presented at Fall Meeting, Power Generation Committee.

A review of the BCR investigation of coal washability at 30 mesh x 0 and p.c. sizes as reported in (JJ)a and (JJ)b, with emphasis on the 29 coals that originated in Pennsylvania out of a total of 90 tested.

(LL) Chemical Construction Corp. (New York, N.Y.) Engineering analysis of emission control technology for sulfuric acid manufacturing processes. National Air Pollution Control Administration Report March 1970, (Contract CPA-22-69-81) in 2 volumes; Vol. 1 PB-193 393, 324 pp., Vol. 2, PB-190 471, 144 pp.

Vol. 1 gives a survey of more than 70 systems and devices for the control of emissions of SO_2, H_2SO_4 mist and nitrogen oxides from sulfuric acid plants. It is concluded that adequate control can be achieved at almost any H_2SO_4 plant from a technical viewpoint, but not always economically. Analyses of 20 of the more advanced and potentially promising systems show the estimated costs of emission control under various plant conditions. It is noted that systems suitable for emission control in power plants are not necessarily suitable and economical for H_2SO_4 plants. Further study of the problems of emission control in H_2SO_4 plants is recommended.

Vol. 2 is a survey of literature giving three bibliographies with abstracts: (2) Removal and recovery of sylfur oxides from H_2SO_4-plant tail gases, 1955-1967; (b) SO_3 and H_2SO_4 mist emissions and their control, 1907-1967; (c) Removal of nitrogen oxides from Chamber and Mills-Packard H_2SO_4 plant tail gases, (1907-1968).

(MM) Stites, J.G., Jr., Horlacher, W.R., Jr. (Monsanto Co., St. Louis, Mo.), Bachofer, J.L., Jr. and Bartman, J.S. (Metropolitan Edison Co., Portland and Reading, resp., Pa.) Removing SO_2 from flue gas. Chem. Eng. Prog. 65, Oct. 1969, p. 74-79.

The Monsanto process for SO_2 removal by catalytic oxidation was put through pilot plant test at Metropolitan Edison

Company's generating station at Portland, Pa. Operation started in August 1967, with a flue gas capacity of 1, 440, 000 SCF/hr. which was about 1/16 of the total flue gas from the 250, 000 kw. No. 2 generating unit. The flue gas was first freed from fly ash and was passed through the catalyst bed at about 650° F whereby the SO_2 in the flue gas was about 90% oxidized to SO_3. The SO_3 combines with the moisture in the flue gas to form H_2SO_4. The effluent gas was cooled by heat exchanger to about 450° F and was scrubbed with cold 80% H_2SO_4 in a packed column maintained at 225° F. After that the gas was passed through a Brink mist eliminator and the residual gas was vented. A slip-stream of the acid from the bottom of the absorption column was taken as the product acid containing 80% H_2SO_4. The SO_2 removal efficiency was 90%, and H_2SO_4 recovery was 99. 5% based on SO_3. The installation cost of the Monsanto system was estimated at $20-$30 per kw power plant capacity. The operating cost would be balanced by the value of recovered H_2SO_4 based on a coal containing 3% S and a 100% H_2SO_4 price of $13. 50/sh. ton.

(NN) Anon., Chemical Week, Vol. 104 No. 4, July 22, 1970. Anon., Chem. and Eng. News, Vol. 48 No. 30, July 20, 1970.

Illinois Power Company filed proposal with Illinois Commerce Commission to install a demonstration unit of Monsanto Envirochem Systems, Inc. Cat-Ox method for removing sulfur oxides and particulates from plant stack emissions at IP's Wood River station No. 4. The Cat-Ox system will recover sulfur as sulfuric acid. The facility is expected to cost $6. 8 million, which will be funded $3. 3 million by IP and $3. 5 million by NAPCA.

(PP) Shah, I. S., The removal and recovery of sulfur dioxide and acid mist from sulfuric acid plant stack gas using Venturi Scrubbers, A. I. Ch. E. -IMIQ 3rd Joint Meeting, Denver, Colorado, August 1970.

Sulfur dioxide, trioxide and sulfuric acid mist are atmospheric pollutants from combustion of sulfur bearing fossil fuels, smelting of sulfide ores, sulfuric acid plants and other processes. The venturi scrubber is capable of high efficiency in treating sulfur oxide and mist containing gases. The magnesium base SO_2 removal and recovery process can limit outlet SO_2 concentrations to 50-250 ppm. Dried magnesium sulfur salts heated in the presence of coke will release SO_2 when directly heated to yield a 15% SO_2 concentration in the combustion gas, and regenerate magnesium oxide for recycle to the scrubbers.

(QQ) Cronan, C.S., Magnesia pulping breaks pollution saltemate. Chemical Engineering, Sept. 8, 1958.

New magnesia-base sulfite pulping on stream at Brown Paper Co., uses $Mg(OH)_2$ liquor in Venturi scrubbers to recover 1-2% SO_2 from flue gas of 160 M lb/hr boiler.

(RR) Anon., Chemical Week, Vol. 104 No. 3, July 15, 1970.

Sulfur dioxide removal system developed by Chemical Construction Corporation and Basic Chemicals Division of Basic, Inc., will be installed at the 150 MW Mystic station of Boston Edison Co. The sulfur dioxide will be converted to magnesium sulfur salts at the Mystic plant and shipped to Essex Chemical Corp. at Rumford, R.I. for regeneration as sulfuric acid and magnesium oxide. The latter will be returned to the power plant for reuse. The cost of the facility is estimated at $5 million to be funded jointly by NAPCA and the other participants.

(SS) Sulphur Patents Limited, Billingham, England, Brochure: The recovery of sulphur from smelter gases, no date, 28 pages, 4 figures.

Absorption and discharge of SO_2 from solutions of basic aluminum sulfate is described, and account of installation at Imatra, Finland.

Also an account of the Boliden process for reduction of SO_2 to sulfur with producer gas from coke. Chemistry is described.

(TT) Watt, S.G., Wellman-Lord SO_2 recovery process, Brochure: Wellman-Lord, Inc., undated, 19 pages, incl. simplified flow sheet.

The Wellman-Lord SO_2 recovery process is based on absorption of SO_2 in potassium sulfite solution, crystallization of $K_2S_2O_5$ from this solution, and conversion of $K_2S_2O_5$ to $KHSO_3$ by dissolving the crystals in water. SO_2 is stripped from the $KHSO_3$ solution and can be used as a gas or compressed for shipment. The 3 year development program included a 1 MW pilot plant at Tampa Electric's Gannon Station and a 25 MW demonstration plant in operation at Baltimore Gas and Electric's Crane Station. Technical and economic performance have been promising. Tests have been conducted at a metallurgical smelter and this process concept is held to have wide application. An alternate sodium sulfite system is available. Pilot plant tests to produce elemental sulfur are in progress.

(UU) Craig, T.L., Hughes, F., Watt, W.G., Commercial experience - Wellman-Lord SO_2 recovery process, Third joint meeting A.I.Ch.E.-IMIQ Denver, Colorado August 30/Sept. 2, 1970.

Brief history of Wellman-Lord development activity with alkali metal SO_2 absorbent systems and preliminary comment on operation of prototype (45,000 SCFM) unit at Paulsboro for control of sulfur oxide emissions from refinery acid regeneration.

(VV) Kohl, A.L. and Riesenfeld, F.C., Gas Purification, McGraw Hill Book Co., Inc., N.Y., 1960; pp 199/210 incl.

Sulfur dioxide recovery processes employing aromatic amines are discussed. Xylidine-water mixtures of the "Sulphidine" process and dimethylaniline as used by Falconbridge Nickel Company and American Smelting and Refining Co. in analogous chemistry based on formation of the sulfites of the aromatic amines and stripping with heat in counterflow exchangers. The xylidine-water absorbent is more efficient at SO_2 gas concentration below about 2%. DMA is competitive or superior at higher concentrations.

(WW) King, R.A. Economic utilization of sulfur dioxide from metallurgical gases, Industrial and Engineering Chemistry, Vol. 42, No. 11 November 1950.

The market oriented history of sulfur dioxide recovery at Trail, B.C. is reviewed. Sulfuric acid, ammonium sulfate and sulfur manufacture is described.

(XX) Fleming, E.P. and Fitt, T.C., Industrial and Engineering Chemistry, Vol. 42, No. 11, November 1950.

Direct reduction of sulfur dioxide in metallurgical gases by reaction at high temperature with methane as practiced by American Smelting and Refining Co.

X

HIGH SULFUR COMBUSTOR STUDY

Review of Literature

SUBJECT INDEX

A

Aerosol emissions from sulfuric acid plants (U 25, 151)
Air classifiers for pyritic sulfur removal from coal (JJ 101)
Air cleaning by wet scrubbing (Z33)
Air cleaning systems cost calculation (W166)
Air filters, wet type (W 70)
Air pollution control credit of low-sulfur coal (F ii-4)
Air pollutant emissions from coal- and oil-fired boilers (Z 25)
Air pollutant emissions from non-ferrous smelting (Z 24)
Air pollutant emissions from sulfuric acid plants (Z 25; AA 79)
Air pollutants produced in U.S., 1965 (S p. ii)
Air pollution control costs (AA 2, 3, 25, 27)
Air pollution control costs to industry (AA 27)
Air pollution control for fluid-bed furnaces (Z 28)
Air pollution control in fluidized-bed boiler plants (M 10)
Air pollution control in power stations, costs (AA 37)
Air pollution control in sulfuric acid plants, costs (AA 79, 81)
Air pollution control laws (U.S.) (A)
Air pollution control regulations (A 27, 43, 44)
Air quality criteria for dust (V 189)
Air quality criteria for particulate pollutants (U 189)
Air quality improvement costs (S 315)
Air quality improvement methods (S 314)
Air quality standards (A 27, 43, 44)
Alkalized Alumina process capital investment (EE 28)
Alkalized Alumina process for SO_2 removal (EE ii, 6, 18; GG 328)
Alkalized Alumina process, operating costs (EE 28, 46)
Alkalized alumina process (X 49)
Alkalized Alumina process for SO_2 removal (S 147)
Alkalized Alumina process for sulfur oxides removal (Z 40, 41)
Alkalized Alumina process - W.R. Grace & Co. report (Z 41)
Alkalized Alumina process - M.W. Kellogg report (Z 41)
Alkalized Alumina process - Oak Ridge National Laboratory report (Z 41)

Alkalized Alumina process operating costs (X 51)
Alkalized Alumina process - U.S. Bureau of Mines reports (Z 4)
Alkalized Alumina process - Avco report (Z 40)
Allied Chemical Corporation (Z 20)
Allied Chemical process for sulfur recovery from roaster gases (R vi-a-16)
Anthracite refuse utilization (H 48)
Asarco process flow diagram (FF f-2; GG 168, 225, 228)
Asarco SO_2 reduction capital costs related to SO_2 concentration (R viii-c-18, Table B-5)
Asarco process operating costs (FF 16; GG 240, 252)
Asarco SO_2 reduction operating costs related to SO_2 concentration (R viii-c-19, Table B-6)
Asarco Process capital investment (FF 15, f-8; GG 176, 241, 253)
Asarco Process for SO_2 reduction (Riii-3, vi-a-11)
Asarco process for SO_2 reduction, flow diagram (R vi-a-12)
Asarco process for sulfur recovery from smelter gases (FF 8, f-2, f-8, f-9; GG 40, 42, 43, 168, 169)
Ash content of U.S. coals (N 5)
Atomics International (molten carbonate) process for SO_2 removal (EE 14)
Averbukh's data on SO_2 reduction by methane (GG 272)

B

B&W-Esso - MgO system for sulfur oxides recovery flow diagram (Y 3)
B&W-Esso - system of sulfur recovery (Y 4, 5)
B&W-Esso - system of sulfur oxides recovery, flow diagram (Y 4)
B&W - MgO system of sulfur recovery (Y 3, 5)
Babcock & Wilcox Co. (Z 10, 22)
Bag filters (W 109)
Battelle Memorial Institute (Z 8, 9, 20, 24, 39, 42, 44)
BCR process for coal gasification (X 116)
Bechtel Corporation (Z 12, 15)
Beckwell process for SO_2 recovery (X 56)
Belco Industrial Equipment Company (Z 34)
Bituminous coal compositions, eastern U.S. (B i-2)
Bituminous Coal Research, Incorporated (Z 15)

Brick made with coal refuse (H 50)
Bricks made of flyash (H 46)
Burning refuse for power generation (Z 35)

C

"Carbonization-desulfurization" process for coal tailiers (B I-3, iv-3)
Catalytic oxidation process for SO_2 recovery (X 54)
Catalytic oxidation process for SO_2 removal (EE ii, 5, 9, 18)
Catalytic oxidation process for SO_2 removal operating costs (X 56)
Centrifugal dust collectors (W 9, 44, 50)
Claus process at low temperature (GG 7, 56, 72, 157, 158, 321)
Claus process at low temperature for sulfur recovery from smelter gases (GG 87)
Claus process at low temperature for stack-gas desulfurization (GG 72)
Claus process at low temperature, reaction equilibrium (GG 90)
Claus process at low temperature, thermodynamics (GG 90)
Claus process for sulfur from hydrogen sulfide (R vi-a-15)
Claus process (low temperature) catalyst adsorption (GG 157)
Claus process (low temperature) catalyst regenerator sketch (GG 196)
Claus process (low temperature) cross flow reactor sketch (GG 282)
Claus process (low temperature) flow diagram (GG 193, 202, 204, 276)
Claus process (low temperature) moving bed reactor (GG 201)
Claus process (low temperature) moving bed reactor, capital investment (GG 280)
Claus process (low temperature) moving bed reactor, operating costs (GG 279)
Claus process (low temperature), operating costs (GG 159, 207, 210, 212, 215, 218, 221, 261, 265, 273)
Claus process (low temperature) switch reactor, detailed sketch (GG 278)
Claus process (low temperature) switch reactor, flow diagram (GG 276)

Claus process (normal) for sulfur from H_2S	(GG 7, 19, 54, 313)
Claus process (normal), operating costs	(GG 153)
Claus process (low temperature), capital investment	(G 208, 211, 213, 216, 219, 222, 263, 274)
Claus process using moving beds	(GG 74, 86)
Coal analysis, U.S.	(D 11, E 14)
Coal ash composition, U.S.	(H 52)
Coal ash fusion temperature	(T 8, 36)
Coal ash utilization	(H 46)
Coal beneficiation by deep cleaning	(B iii-5)
Coal burning in fluidized bed	(G 52)
Coal burning, powdered coal	(G 51)
Coal cleaning by dense medium cones	(G 40)
Coal cleaning by froth flotation	(G 40)
Coal cleaning data on U.S. coals	(JJ 181)
Coal cleaning equipment study	(Z 14)
Coal cleaning methods	(KK 5)
Coal cleaning potentials	(X 36)
Coal cleaning plant cost analysis	(Z 16)
Coal cleaning processes	(Z 13)
Coal composition, U.S.	(B A-2; E 14)
Coal composition, U.S. coals	(Z 15)
Coal consumption by power plants in U.S.	(AA 42)
Coal consumption by power plants, U.S. 1967, by state	(K 49, 51)
Coal consumption in U.S.	(CC f-1)
Coal consumption in U.S. by regions	(CC t-1)
Coal consumption, U.S.	(D 8, 10; E 8, 10)
Coal, deep-cleaned, composition	(B A-2)
Coal desulfurization	(S 91, 94, 101, 104)
Coal desulfurization	(X 34, 109, 117)
Coal desulfurization by froth flotation	(Z 15)
Coal desulfurization processes compared	(Z 14)
Coal fields, U.S.	(X 10)
Coal float-sink tests	(CC 5, f-1, f-5)
Coal heating values, U.S.	(D 11; E 14)
Coal mines in Illinois	(Z 15)
Coal mining companies, U.S.	(J 7)
Coal mining flow diagram	(H - inside cover)
Coal preparation	(H - inside cover)
Coal preparation, commercial practice	(B iii-4)
Coal price appreciation due to sulfur removal	(F ii-2)
Coal processing for sulfur removal	(E 48; F iii-20)
Coal pulverization	(G 38)

Coal pyrites, grades and analysis (B i-2, iii-6, vii-9 app. A-2; F iii-50)
Coal pyrites processing feasibility (Z 14)
Coal refuse (See also under coal tailings)
Coal refuse disposal (X 91)
Coal reserves, U.S. (X 11)
Coal residue sulfur content, Illinois (T 9, 38)
Coal tailings composition (B i-2, iii-6, vii-19-20, app. A-1-2; F iii-50)
Coal tailings, recoverable values (B i-13ff, vii-21, vii-38, vii-40)
Coal tailings roasting in fluidized bed (B v-2)
Coal up-grading by washing (T 9, 38)
Coal washability (Z 13)
Coal washing (CC 4, f-4, f-g, t-2, t-3; DD 8; G 38, 40)
Coal washing, capital investment (CC t-4, t-5)
Coal washability data (E 62, 75, 82, 89, 92, 100, 107)
Coal washability graphs (E 66, 76, 81, 83, 88, 90, 94, 99, 101, 106, 108)
Coal washing for sulfur removal (D 36)
Coal washing operating costs (CC t-4, t-5)
Coal washing plant flow diagrams (E 71, 77, 84, 91, 102, 109, 112, 114, 116, 128; X 35, 36, 38)
Coal washing vs. specific gravity (CC f-2, f-3, f-4)
Coals in Appalachian region (DD 4)
Coals in U.S. by rank (N 4, 5)
Coals, U.S., ash and sulfur content (CC t-3)
Combustion emissions control (Z 26)
"Combustion-SO_2 removal" process for coal tailings (B i-3ff, v-12)
Cominco process flow diagram (R vi-a-9)
Cominco process for SO_2 recovery, capital investment (FF 13, f-5)
Cominco process for SO_2 recovery, flow diagram (FF f-1)
Cominco process for SO_2 recovery from smelter gases (FF 7, f-1)
Cominco process for SO_2 recovery operating costs (FF 14, f-6, f-7)
Cominco process for up-grading SO_2 gases (R iii-2, via-8)

Commercial Testing and Engineering Company (Z 15)
Compound water cyclone for pyritic sulfur removal from coal (JJ 6, 69, 73)
Concentrating spiral for pyritic sulfur removal from coal (JJ 6, 86, 91)
Concentrating table for pyritic sulfur removal from coal (JJ 6, 35, 37)
Cyclones for dust removal (W 9, 44, 50)

D

DAP-Manganese process for SO_2 removal (S 148)
DAP-Manganese process for SO_2 removal (EE 6)
Dimethyalaniline (DMA) process for SO_2 recovery (R iii-3, vi-a-1, vi-a-7)
Dust removal equipment (wet type) costs (W 158)
Dust removal equipment efficiencies (W 155)
Dust removal from gases (S 136)
Dust removal from flue gases, costs (S 138)
Dust removers, centrifugal type (W 44)
Dust removers, venturi type (W 58)
Dust removers, wet types (W 50)

E

Economic gains through coal desulfurization (F ii-22, iii-15)
Electrostatic precipitation high-voltage type (W 81)
Electrostatic precipitation low-voltage type (W 96)
Electrostatic precipitators (Z 33, 34)
Electrostatic precipitators for high-resistivity dust (Z 34)
Energy consumption, U.S., 1950-1955 (L 13)
Energy cost calculation (BB 13 t-3)

F

Float-sink test data on U.S. coals (JJ 111)
Float-sink testing of coals (Z 15)
Float-sink tests for coal (CC 5, f-1, f-5)
Fluidized-bed coal burning boilers (M 1)
Fluidized-bed combustion (Z 27)
Fluidized-bed fuel diluent material (M 3)
"Fluidized-bed roasting" process for pyrites (B i-7)
Flyash bricks (H 46)
Flyash removal (S 139)
Flyash utilization (S 141)

G

Gravity separation for pyrite removal from coal (B iii-3)
Grill (Mn-Mg Oxide) process for SO_2 removal (EE 11)
Grillo process for SO_2 removal from flue gas (X 57)

H

Hardgrove grindability, Illinois coal (T 9, 37)
Heating required for making H_2SO_4 from low-SO_2 gases (R vi-a-5)
High-sulfur coals utilization with coal beneficiation (B i-11)
High-sulfur combustor study (Z 16)
Hitachi (active carbon) process for SO_2 removal (EE 10)
Humprey spiral for pyrite removal from coal (Z 14)
Hydrogen sulfide from reformer natural gas - Sulfur Reaction (GG 199, 202)
Hydrogen sulfide - reformed natural gas reaction equilibrium (GG 200)
Hydrogen Sulfide synthesis from SO_2 and methane (GG 131, 198)

I

Illinois coal ash fusion temperature (T 8, 9, 36)
Illinois coal sulfur content (T 5, 11, 31, 33, 34, 38)
Illinois Geological Survey (Z 9, 15)
Iron from coal tailings (B iii-8)
Iron oxide content in coal tailings (B vii-20)
Iron oxide reduction in pyrite cinders (B v-7)
Iron oxide (Siemens Schuckerwerk, Germany) process for SO_2 removal (EE 12)
Iron recovery from coal pyrites (C 3, 10, 39; F iii-100, iii-110)
Iron recovery from pyrite cinders, process flow diagram (C 23)
Iron recovery from coal pyrite, process flow diagram (F iii-107)

K

Kiyoura-TIT (catalytic oxidation-NH_3 injection) process for SO_2 removal (EE 9)
Kiyoura-TIT process for SO_2 removal (S 150; EE 9)

Kulscar process, capital investment (GG 178, 259)
Kulscar process, flow diagram (GG 172)
Kulscar process for SO_2 reduction (GG 37, 46, 172, 173, 178, 183, 271)
Kulscar process operating costs (GG 184, 258)

L

Low-sulfur coal reserves, U.S. (S 51)
Lurgi process for SO_2 removal from flue gases (X 57)

M

Arthur G. McKee & Company (Z 24)
Mitsubishi (Red Mud) process for SO_2 removal (EE 13)
Molten carbonate process for SO_2 removal from flue gases (X 57)
Molten salt process for coal gasification (X 116)
Monsanto (catalytic oxidation) process for SO_2 removal (EE 5)
Monsanto process for SO_2 recovery (B i-13; EE 5)
Municipal air pollution control laws (S 176, 183, 332)

O

Organic sulfur in Illinois coal (T 34)
Organic sulfur in U.S. coals (D 11; E 27; JJ 111)
Outokumpu process for high-sulfur coal (B v-22)
Outokumpu process for pyrites (B i-9, v-14)

P

Particle-size effect on coal washing (DD 8)
Particulate measurements for combustion products (Z 37)
Particulate pollutant emissions from burning coal (W 3, 7, 150)
Pope, Evans and Robbins, Incorporated (Z 28)
Power Plant costs based on burning coal refuse (B app. D-1)
Power plant operating costs, U.S., 1955-1968 (L 26)
Princeton Chemical (SO_2 reduction) process for SO_2 removal (EE 13)
Pyrite beneficiation (B iii-5; C 15, 39; F iii-20)
Pyrite cinders composition (F iii-101)
Pyrite concentration (B iii-5; C 15, 39; F iii-20)

Pyrite content in coal refuse (C 3)
Pyrite from coal refuse, analysis (C 6, 7)
Pyrite recovery from coal (C 15, 39; F iii-20)
Pyrite removal from coal (KK 1)
Pyrite removal from fine coal (KK 4)
Pyrite removal methods for fine coals (Z 15)
Pyrites in coal beds (KK 1)
Pyrites in Illinois coal (T 34)
Pyrites removal from coal by tabling (Z 14)
Pyrites removal from coal (Z 13)
Pyrites removal from coal by kinetic separation (Z 15)
Pyritic sulfur in U.S. coals (JJ 111; KK 6; D 11; E 27; N 5)
Pyritic sulfur fremoval from U.S. coals (Z 14, 15)
Pyritic sulfur removal from coal (X 35)
Pyrites removal from U.S. coals (Z 14, 15)
Pyritic coal utilization economics (Z 15, 16)

R

Red Mud (Mitsubishi) process for SO_2 removal (EE 13)
Reinluft (carbon adsorption) process for SO_2 removal (EE ii, 6, 23)
Reinluft process capital investment (EE 28, 48)
Reinluft process for SO_2 removal from flue gases (X 56)
Reinluft process operating costs (EE 28, 49, 50)
Roberts and Schaefer Company (Z 16)

S

Showa Denko (ammonia scrubbing) process for SO_2 removal (EE 9)
Spiral concentrator for pyritic sulfur removal from coal (JJ 6, 86, 91)
Stack costs (X 101)
Stack location evaluations (X 99)
State air pollution control laws (S 175, 179)
Steam generator emission control by alkaline solution scrubbing (Y 5)
Steam generator, pulverized coal fired (BB f-8, f-12, f-14, f-15; f-16, f-18)
Steam generators, capital costs (BB t-2, f-20)
Steam generator emission control by MgO slurry (Y 3, 5)
Steam power plants, capital costs (BB 11, t-2)
Steam power plants, fuel costs (BB 13, t-3)

Steam power plants, operating costs	(BB 13, t-3)
Steam turbines - electric generators, capital costs	(BB t-2, f-21)
Stone & Webster-Ionic (sodium sulfite) process for SO_2 removal	(EE 14)
Sulfacid process for SO_2 removal	(LL I-iv-37, I-iv-f-21)
Sulfur consumption, U.S., 1945-1970	(EE 32)
Sulfur content in coal refuse	(D 48, 53; E 48; F iii-20)
Sulfur dioxide content in H_2SO_4 plant tail gases	(R vi-a-5)
Sulfur dioxide content of gas from burning coal pyrite	(F iii-71)
Sulfur dioxide gas up-graded by DMA process	(R iii-3)
Sulfur dioxide, liquefied market	(R iii-4)
Sulfur dioxide-methane reaction equilibrium	(GG 131, 165, 185, 268)
Sulfur dioxide minimum concentration required for economic H_2SO_4 production	(R vi-a-4)
Sulfur dioxide recovery by ammonium sulfite solutions	(FF 7, f-1)
Sulfur dioxide recovery from smelter gases, Cominco process	(FF 6, f-1)
Sulfur dioxide recovery systems for H_2SO_4 plant tail gases	(LL I-iv-60)
Sulfur dioxide reduction, Asarco process costs	(R iii-3, vi-a-11)
Sulfur dioxide reduction by carbon, thermodynamics	(GG 13)
Sulfur dioxide reduction by coke	(GG 6, 13, 110, 111)
Sulfur dioxide reduction by coke, capital investment	(GG 120, 124, 127, 130)
Sulfur dioxide reduction by coke, operating costs	(GG 122, 126, 128, 156, 159)
Sulfur dioxide reduction by hydrocarbons	(GG 19, 24, 25, 309)
Sulfur dioxide reduction by H_2-CO mixed gases	(GG 7, 290)
Sulfur dioxide reduction by H_2S, flow diagram	(GG 202)
Sulfur dioxide reduction by methane	(FF 10, f-2; GG 6, 19, 24, 25)
Sulfur dioxide reduction by methane, flow diagram	(GG 225, 228, 229)
Sulfur dioxide reduction by methane, Yushkevich process	(GG 162, 163)
Sulfur dioxide reduction by reformed natural gas	(GG 7, 48, 50)
Sulfur dioxide reduction costs	(GG 7)
Sulfur dioxide removal by B&W-Esso process	(Y 5)
Sulfur dioxide removal by moltan alkali carbonates	(Z 21)
Sulfur dioxide removal by Na_2SO_3-ZnO process	(R vi-a-22)
Sulfur from hydrogen sulfide, Claus process	(GG 7, 19, 54, 56, 72, 313, 321)

Sulfur from SO_2, Asarco process	(FF 8, f-2; GG 40, 168, 183)
Sulfur from sulfur dioxide by reduction, bibliography	(GG 283)
Sulfur oxide emissions from burning coal	(X 9)
Sulfur oxides removal from stack gases, alpha-Fe_2O_3 process	(EE 12)
Sulfur oxides removal from stack gases	(EE ii, 4)
Sulfur oxides removal from stack gases, active carbon (Hitachi) process	(EE 10)
Sulfur oxides removal from stack gases, Alkalized Alumina process	(EE ii, 6, 8, 21)
Sulfur oxides removal from stack gases by manganese hydroxide (Mitsubishi)	(EE 14)
Sulfur oxides removal from stack gases by molten carbonate	(EE 14)
Sulfur oxides removal from stack gases, Chevron process	(EE 13)
Sulfur oxides removal from stack gases, Kiyoura-TIT process	(EE 9)
Sulfur oxides removal from stack gases, Mn-Mg oxide (Grillo) process	(EE 11)
Sulfur oxides removal from stack gases, Monsanto process	(EE 5)
Sulfur oxides removal from stack gases, Showa Denko process	(EE 9)
Sulfur oxides removal from stack gases by sodium sulfite solutions	(EE 14)
Sulfur oxides removal from stack gases, UOP process	(EE 5)
Sulfur oxides removal from stack gases, Wellman Lord process	(EE 7)
Sulfur oxides removal from stack gas, Wickert process	(EE 9)
Sulfur oxides removal processes, operating costs	(EE 28)
Sulfur prices forecast to 1975	(R iii-4)
Sulfur recovery, Claus process	(X 78, 79)
Sulfur recovery, Claus process, costs	(X 81)
Sulfur recovery from smelter gases, Texas Gulf Sulfur process	(FF 10)
Sulfur removal from coal, costs	(CC f-8)
Sulfuric acid from coal pyrites, material balance	(F iii-54)
Sulfuric acid from coal pyrites, process flow diagram	(F iii-43)

T

Texas Gulf Sulfur (TGS) process for SO_2 reduction (FF 10; GG 34, 43, 46)

W

Washability test data on U.S. coals (JJ 18, 25, 181)
Paul Weir Company (Z 14)
Wellman-Lord process for SO_2 removal (S 154)
Wellman-Lord process (sodium sulfite) for SO_2 removal (EE 7)
West (T GS) process for SO_2 reduction (GG 34, 46, 164, 170)

Y

Yushkevich process for sulfur recovery from SO_2 (GG 162, 163)